A METHODOLOGY FOR THE DESIGN AND EVALUATION OF MINERALS EXTRACTION PROCESSES

by

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ABSTRACT

A methodology has been proposed for the design and evaluation of minerals extraction processes. It consists of combining mass and energy balance, and capital and operating cost estimation models of process unit operations to develop an overall process simulation; this can then be used to provide technical and economic analyses of the process.

An extensive analysis of the literature and industrial practice concerning process design, unit operation modelling, process simulation and costing and economic evaluation has been carried out and recommendations made for the computer-based implementation of the methodology.

Initial development work on the feasibility study application of the methodology has been based on a state-of-the-art flowsheeting package, ASPEN PLUS. Additions were made to the capabilities of this system to enable it to simulate a variety of minerals extraction processes and a separate process costing and economic evaluation system was developed to initiate these aspects of the methodology.

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A Methodology for the Computer-Aided Design and Evaluation of Minerals Extraction Processes

1.0 Introduction

Engineering design can be said to be a process of planning and decision making to produce information to ensure the correct output under optimal conditions; there is therefore a requirement for a major effort in the synthesis and evaluation of alternative methods of implementing an idea. Each alternative implies a variety of consequences, some of which may be predictable, uncertain or unobservable according to the different input and internal variables involved; thus the aim of design and decision making is to choose the optimal configuration (1).

In the mineral industries the formula for achieving this configuration has undergone significant changes during history. Initially mineral processing could be regarded as labour intensive, but increasing demand, whilst variously benefitting locally from cheap labour, energy and/or capital, has resulted in a gradual change to mechanization and expansion of production. In the 1970's and 1980's however, rising world consumption has meant that the achievement of low production costs are increasingly dependent on a careful balance of expensive labour, expensive energy and expensive capital (2, 3).

Thus, coupled with the increasing size and complexity of mining ventures and the multiplicity of factors to be considered, the task of choosing the optimal process configuration has been made more and more demanding. The ability to engineer more efficient processes has been limited by the large effort required for process engineers to evaluate alternative flowsheets (4).

In recent years the need for increased sophistication in the design of modern metallurgical installations has become pronounced, as has the need for techniques to permit process engineers to evaluate many more process configurations than is now possible. Many of the traditional methods of plant design have been found inadequate to meet current industrial needs (5). For many years the slide rule and the desk-top mechanical calculator were the mainstays of the process

engineer and often rigorous solutions to complex problems were avoided, compromise dictating an approximation method with an unquantified risk factor (6). Nowadays engineers are turning to the use of computers with increasing frequency, the advent of efficient computing equipment permitting more detailed analyses to be carried out with a reasonable degree of effort and cost.

However, despite considerable development and investment in computer technology within many areas of the mining industry, the use of computer aided process design and evaluation lags far behind many other process industries.

1.1 Mineral Process Evaluation

The final product of a process evaluation is an estimate of the potential economic viability, taking into consideration all costs and revenues over the project life. The methods used to carry out an evaluation within the mineral industry are many and varied, and are generally dependent on the customs and requirements of the body for whom the study is being carried out. However, the general approach to the development of a new process plant is fairly standard (7 - 9) and the steps involved in any evaluation are common to all.

The requirements of an economic feasibility study have been fairly well summarized in a description of the process evaluation office of the U.S. Bureau of Mines (10):-

1) Background data from research and literature
2) Preparation of material and energy balance
3) Equipment layout and sizing
4) Costing equipment
5) Estimation of capital and operating costs
6) Calculation of cash flows, rates of return, payback times, etc.
7) Evaluating effects of modifications to circuit or changes in costs.

Using traditional means, many of these steps are time consuming and repetitive; one design company estimates that the mass balance calculations for just one flowsheet can take up to 50 man hours (11). Therefore, to increase the speed of evaluations, and hence the possible depth of analysis, some automation is, at least, desirable.

Steps 2 - 7 of the USBM evaluation have been adapted to computer methods within the chemical industry and the use of integrated systems combining the individual components into a complex evaluation is widespread. However, despite similar automation of some of the evaluation components, there has been little or no work on developing similar systems for the minerals industry.

1.1.1 Material and Energy Balance Calculations

Due to the unique properties of any ore, in metallurgical plant design the preparation of mass and energy balances always involves the use of small scale testwork to determine the physical and chemical properties of the raw material; this information is then scaled up to predict the performance of the industrial scale process. Thus a major part of plant design is the acquisition and extrapolation of data from small scale tests to permit the sizing of equipment for large scale operations; implicit in such an extrapolation is a selection process to allow qualitative or empirical judgements to be made on the likely performance characteristics of specific types of large scale equipment under proposed operating conditions.

There are currently two basic approaches to the acquisition of such data for scale-up; semi-continuous or continuous piloting and prediction from laboratory scale tests, usually of a batch nature.

Piloting is almost certainly the most reliable method providing most of the required parameters for a given circuit except for scale-up factors, and even these are likely to be relatively objective. However, pilot plants are expensive and need large quantities of equipment, time and sample and so are rarely used before the latter stages of development; they also tend to be fairly inflexible when testing alternative process configurations and allow only a limited range of variables to be studied.

It is, thus, often more convenient and economical to use laboratory batch tests to gain the information required for the design of a plant. They generally require a relatively small amount of sample and equipment, and, if well handled, can often produce a great deal of data on the many variables involved. However, such tests cannot describe continuous operation directly; results must be interpreted and transformed, on the basis of empirical correlations or through the use of conceptual models of the processes occurring, to predict

continuous steady state plant behaviour.

Models of many of the unit operations involved in mineral processing plants have been developed during recent years and some of them reviewed in the literature (12 - 15). However, in terms of the development of programs to combine these unit operation models to simulate a complete plant the minerals industry lags behind the chemical industry by some 15 - 20 years (16); this is despite the use of programs to calculate mass balances based on the a priori knowledge of equipment performance (17, 18). It is only recently that programs designed to perform metallurgical plant simulations based on unit operation models have been developed (5, 11, 19-21) and none of these are yet being used in predictive manner, fully developed, publicly released or as sophisticated as their chemical engineering counterparts. The reasons that the mineral industry lags behind in this field are that the computer software capable of handling typical solid-liquid systems was not available until very recently (4, 22, 23); the timespan and cost in developing such software; the modelling of the processes involved in metals extraction is most complex due to the large number of important variables (12); and the empirical nature of the understanding of the fundamentals of the processes.

1.1.2 Plant Cost and Economic Evaluation Calculations

Despite the advantages of process simulators in the evaluation procedure, potentially greater improvements in speed and accuracy are likely to arise from the use of a structured, computerized system for the storage and retrieval of cost information. Not only would this mean an accumulation and pooling of cost data and knowledge over a number of projects, but also it would mean that cost estimates were based on clearly defined data and are likely to take into account all important factors. Thus the two most common and serious errors in costing, those of omission and the use of inapplicable data, are eradicated once the task of setting up the data base and estimation procedure has been successfully achieved. Two such systems have been developed specifically for the minerals industry (24, 25) and are in widespread use.

Even greater advantages would accrue if such a costing system was twinned with a similar financial analysis system. The calculations

to be made by the financial analyst are repetitive and tedious, and highly suited to automation; also the engineer carrying out a process evaluation is unlikely to be experienced in financial matters and his expertise in this field will usually be severely limited (26).

1.2 A Methodology for the Evaluation of Minerals Extraction Processes

Despite the lack of development in some areas of process design and evaluation, and the lack of integration of others, the potential advantages of a computer-aided methodology are so great as to encourage at least an in-depth study of the components and the initiation of a development programme.

The initial step in the methodology proposed in this thesis, is to break down the process flowsheet into its constituent unit operations. Mathematical models, based on the principles of unit operations, can then be developed to relate the values of the input and output process variables (e.g. flowrate, temperature, pressure, solution concentration) for each of these units. These unit models can then be assembled, on the basis of the processing sequence and arrangement among the units, to create an overall model of the process. This model is used to simulate the process and obtain output values at desired input and operating levels. Further models are also developed to relate process variables with the expected values for plant capital costs and operating costs. Hence it is possible to develop annual cash flow estimates for erecting, commissioning and operating the process. These cash flow estimates can be used in a financial evaluation utilizing discounted cash flow techniques, taking into consideration the effects of inflation, taxation and project lifetimes. Sensitivity analysis can also be carried out to identify the cost and energy sensitive areas.

Fig. 1.1 (4) summarizes the steps involved from conception of a new process to start up and operation, and highlights those areas in which a computer-based system can assist. It should be noted that, especially in minerals extraction, such a system can only assist the process designer in understanding and developing the process, and should never totally replace the human input to the design procedure. The methodology, and the systems upon which it is based, proposed in this thesis are intended to replace, or rather supplement, the largely manual methods used presently in the minerals extraction industry.

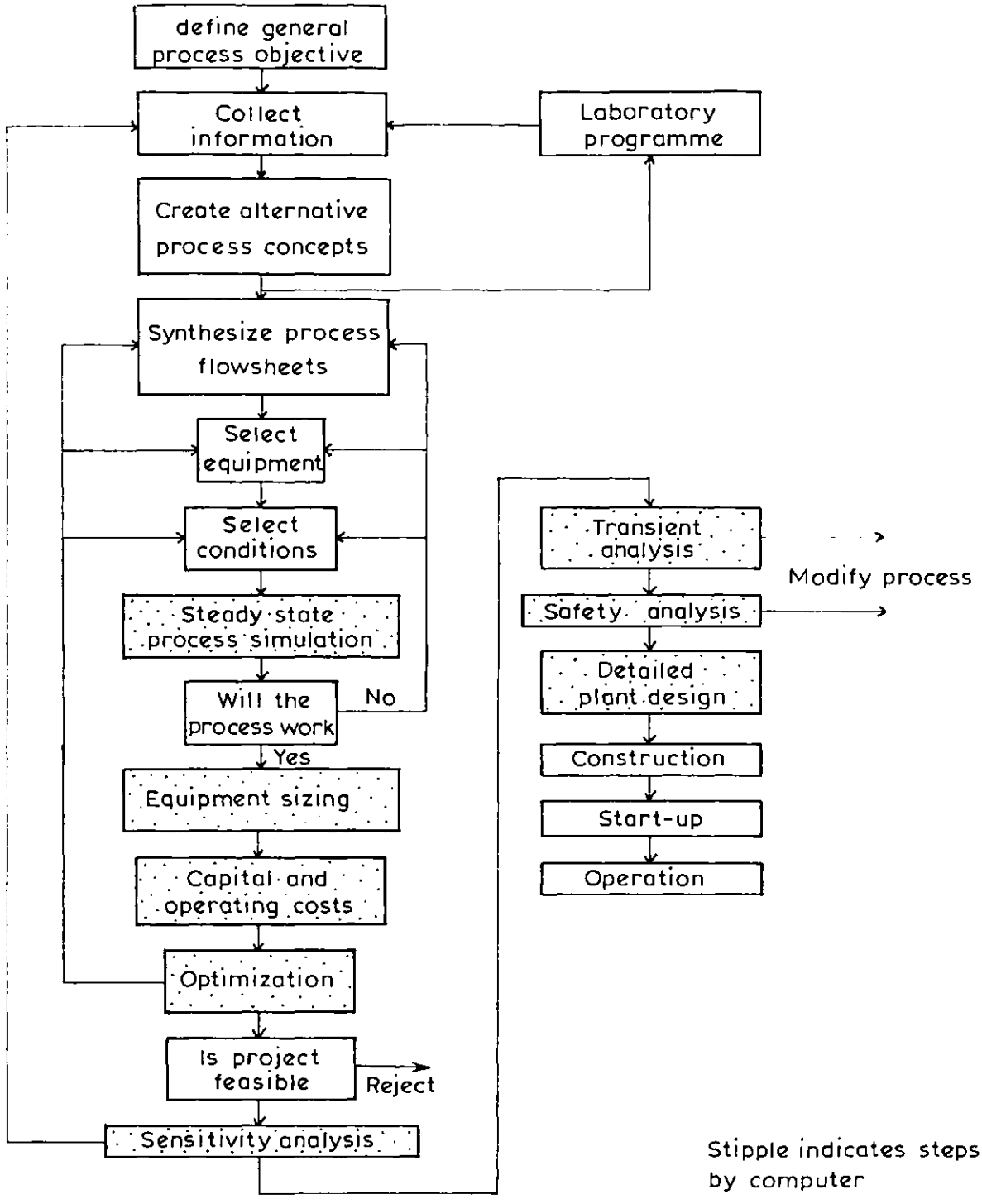


Figure 1.1 Steps in the development of a new process (4)

The advantages of such a methodology include:-

* release of engineers from routine calculations for more creative work.
* improves the number of process configurations that can be studied
* increases the depth to which they can be analysed and hence the understanding thereof.
* increase the number of variables whose effect can be thoroughly studied.
* the methodology is in a modular form and so pools the knowledge and experience of various specialists; it therefore also improves communications and interaction between them.
* improves the understanding of the interaction of project variables and their sensitivity.
* can determine the effects of modifications on operating plants without disrupting production.
* helps to impose some discipline to the evaluation, especially in the costing functions; it thus improves accuracy by eliminating omissions and erroneous shortcut methods.
* cost and economic factors are better emphasised and considered.

Whilst many of the above may appear intangible, in 1975 the Monsanto chemical company estimated the use of a simulation program to save 28 man-months in one project and to derive a potential annual gain of US$50,000 in increased production in another (27). In the minerals industry, even the use of a relatively crude mass balance program (19) was shown to accrue large savings in direct comparison to manual methods.

1.3 Development of a Minerals Extraction Analysis System

In the development of the computer system on which the methodology is based it was necessary to define a strategy such that an attainable objective could be set and reached. If such a strategy had not been followed it could have proved impracticable to demonstrate the potential of the methodology within reasonable time limits.

The strategy followed, paraphrased from Crowe et al (28) and Mular and Herbst (12), was:-

(1) Define the accuracy and generality required and the application of the methodology.

(2) Select modelling techniques for unit operaitons (Chapter 2).

(3) Select those process units to be modelled.

(4) Develop valid models (Chapter 2).

(5) Select plant costing and economic evaluation techniques (Chapter 3).

(6) Select an executive structure and incorporate (4) + (5) as subroutines (Chapters 4 + 5).

(7) Test the methodology and modify as necessary (Chapters 6 + 7).

This strategy was found to be reasonably successful and could be followed if similar systems were to be developed for other applications of the methodology.

As was stated earlier, the steps involved in any process evaluation are basically the same, differing only in the accuracy and included detail. Similarly, the principles of the proposed methodology are sufficiently general as to allow its application to most steps in the exploration and exploitation of a new mineral resource. The various stages where it could prove useful include:-

a) Research and Development: all proposed new extraction processes need to be screened for potential profitability; simple models based on expected performance could be coupled with similar cost estimation models.

b) Exploration stages: in the early stages of developing a new deposit, decisions will need to be made as to whether to carry on project development despite a lack of data; similar models to those used in (a) could be used.

c) Feasibility stage: once more data is available, different process routes will have to be studied and compared to determine the most financially attractive; also different plant sizes and layouts and the effect of changing operating conditions will need to be considered.

d) Detailed plant design: having determined a basic process configuration, detailed design and optimization of each unit operation will need to be carried out; more detailed testwork will be carried out and efficient use made of this extra data.

e) Optimization of operating plant: once the plant has been

designed and commissioned, further analysis and evaluation will still be needed; applications of the methodology include determining the effect of differing raw material, modifications and/or expansion of the existing plant, designing control systems and training plant operators.

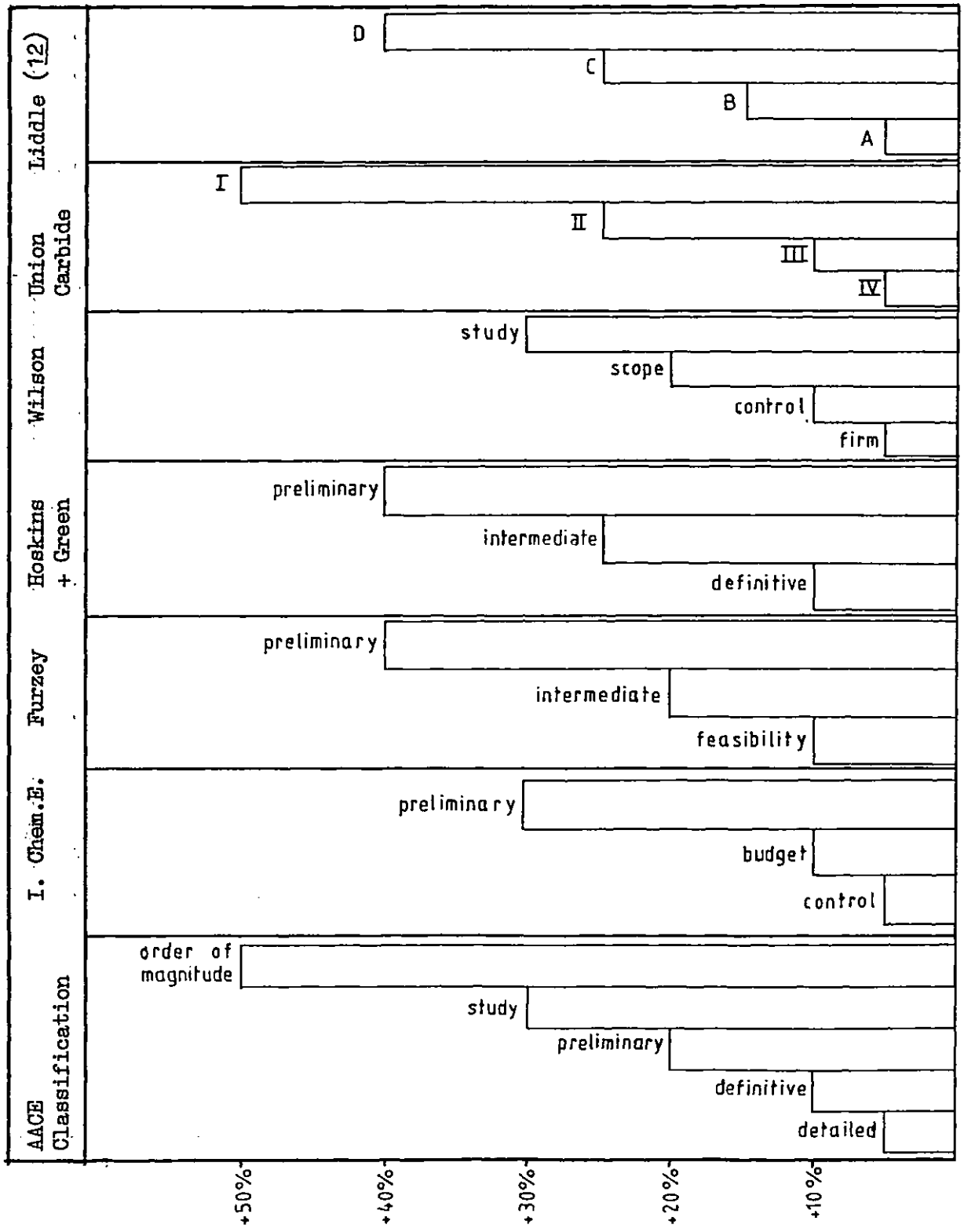
In a study of the literature one will find a proliferation of process evaluation methods and their related, claimed accuracy, and a parallel variety of spurious and confusing names; some of those found in the costing field are shown in Fig. 1.2. A number of authors (29 - 31) have bemoaned this confusion, even criticizing the 'standard definitions' of the American Association of Cost Engineers due to the potentially misleading nomenclature (30); however Liddle's proposed scheme (33) of types A - D appears to be gaining popularity (30) and is a logical route out of the confusion. The type of evaluation produced by a process analysis will largely depend on company philosophy and requirements, the data available and the speed with which it is needed.

Whilst the proposed methodology is applicable to any of the above development stages, each application could require a different supporting computer system and so it is necessary to concentrate on only one. As in chemical engineering and cost estimation (29), it can be argued that the advantages of such a methodology may be most fully exploited and appreciated in the predesign phase of a project; this would relate to Liddle's type C evaluation with maximum errors of approximately +25% to -17%. Thus the system could be used to choose between processing schemes and carry out initial designs, as well as rapidly evaluating the effect of largely external factors associated with the mine planning (e.g. ore grades, tonnages, project life).

From published metallurgical test programme outlines (7-9, 34, 35), it is apparent that this level of analysis can involve up to 2 years' work, and is based on limited ore samples and therefore relatively limited testwork. Extensive testwork may have been carried out in laboratory scale equipment but it is very unlikely that there would be much pilot-scale work. Thus the unit operation model parameters should be obtainable from such tests and should not need to be determined from operating plants.

As the proposed methodology is of general applicability, it is

Fig. 1.2: Estimating levels and nomenclature (after Liddle (33))



desirable that the computer systems upon which it is based should similarly not be process specific. Thus it should be possible to evaluate chemical extraction as well as physical separation processes and also, if possible, pyrometallurgical plants. However, it was decided to demonstrate the methodology using a restricted set of unit operations and to evaluate and show its potential versatility by developing mass balance only models of a wider range of unit operations. As first examples, the processing of coal and low-grade copper ores were initially studied, though the models developed were to be of as general a form as possible.

CHAPTER 2

Unit Operation Modelling

2.0.1 Introduction

In mineral processing the exploration of a system's response to changes in conditions has traditionally been by experimentation. This can be very time consuming, may not allow for a complete understanding of the process and usually results in the study of a narrow set and range of variables. However, it would appear to be the most trusted method of studying a mineral process, far more so than by the use of mathematical models; in most mineral operations the number of factors is too great, the physico-chemical processes too complex and the nature of the raw materials too variable to permit the use of mathematical models for the complete definition of the course of reactions (12). Also, the mineral processing engineer has been comparatively slow to realize the benefits of formal quantitative descriptions of the phenomena that underlie extraction processes.

However the advantages of mathematical modelling, aside from process simulation, are such that it is worthwhile attempting to develop and use suitable models. Once developed they can be used to describe the important characteristics of a system, they enhance the engineer's ability to derive information from data, and can be used to study the effects of different conditions and carry out sensitivity analyses (36, 37).

Due to the complexity of the real process and the limitations of the available mathematics, whatever models used are bound to be fairly idealized and generally give a faithful representation of only a few of the properties of the process. Almost certainly the first models developed are likely to be inaccurate and several attempts will be needed to find a satisfactory model to represent the variables of interest (38); thus the development and testing of models can be a long and expensive procedure. When considering the development of process models for an evaluation system, it is essential that the models used should be tested and proven valid, otherwise an evaluation based on them will be worthless (12). Experience has shown that models,

and computer programs based on them, can be rendered robust and reliable only through extended use by knowledgeable investigators; it is therefore best to upgrade and expand available models rather than to attempt to generate new ones (12, 38).

In developing or selecting a model to describe a system, one must first ask what questions are to be answered by it and what the purpose of the simulation is; this ensures that the model is of the correct form and also governs the selection of process units to be modelled. Thus a steady state simulation does not require time delays, pump boxes, conveyors etc. to be modelled. The basic requirement of a unit operation simulation model is that, given the system inputs and important physical characteristics, the output stream conditions can be calculated.

When selecting and/or developing a model, there are certain important characteristics which should be sought after (36 - 38):-

1) Reliability: the model should be based on theories which have gained general acceptance and which have been well tested.
2) Generality: many models may be useful for a certain purpose or for a narrow range of variables, but are of restricted applicability.
3) Flexibility: the model should be flexible enough to allow alteration of its parameters through a reasonable range of values.
4) Suitability for computing: it may be necessary to judge a model on such factors as running time and memory storage, rather than merely on accuracy.
5) Parameter determination: few models of mineral extraction processes are valid without some corroboratory testwork; the models used in this methodology should have their parameter estimation designed to provide sufficient accuracy with minimum effort.

Mineral treatment processes generally have such a high dimensionality of state that a complete model cannot be formulated. Lynch and Elber (14) have estimated that about 200 parameters are needed to characterise a few essential variables to describe a single flotation cell with moderate similitude. Thus, all that can generally be hoped for is the derivation of a partial model of a process, of a much lower dimensionality of state than the process, and that this model accurately represents the process.

Much excellent work has been and is being carried out to refine the understanding of the phenomena underlying minerals extraction processes and sufficient general knowledge is now available in the literature for most processes, to be confident that progress in modelling will not be significantly impeded by a lack of basic process knowledge. However, theoretical models, fully describing a process without parameter determination, belong to the distant future, as the laws governing processes are too intricate and the number of variables are too numerous.

Mular and Herbst (12) have described the development and verification of empirical models to describe process behaviour. Whilst these are far easier to develop, they offer little insight into the mechanics of the process and can only be used over a narrow range.

Models used in practice in mineral processing are frequently the result of a combination of theoretical and empirical approaches. These "phenomenological" models have a theoretical form which separates the contribution of some "lump" process variables to the process performance. They contain adjustable coefficients that must be calibrated against experimental data collected under appropriate conditions. Even many of the nominally "theoretical" models of mineral processes are phenomenological in nature (12, 37, 39).

Fig. 2.1 illustrates the differences between the three main types of models, though it should be remembered that, in reality, there are no clear boundaries. Whilst restricting the advantages of process simulation it is recognized that theoretical models requiring no parameter determination are a thing of the future and that the best models will usually be of a phenomenological form.

Another type of modelling and simulation, though not strictly outside the classification used above, is linear programming. This has been used quite extensively and successfully in the minerals industry (40, 41), but was considered too limiting and inaccurate to be of wide application in this application.

Finally, although process modelling can be very advantageous when put to optimum use, it also has many limitations that must be recognized. Any model can only be as good as the data and observations on which it was based, and even then few are adequate for all conditions;

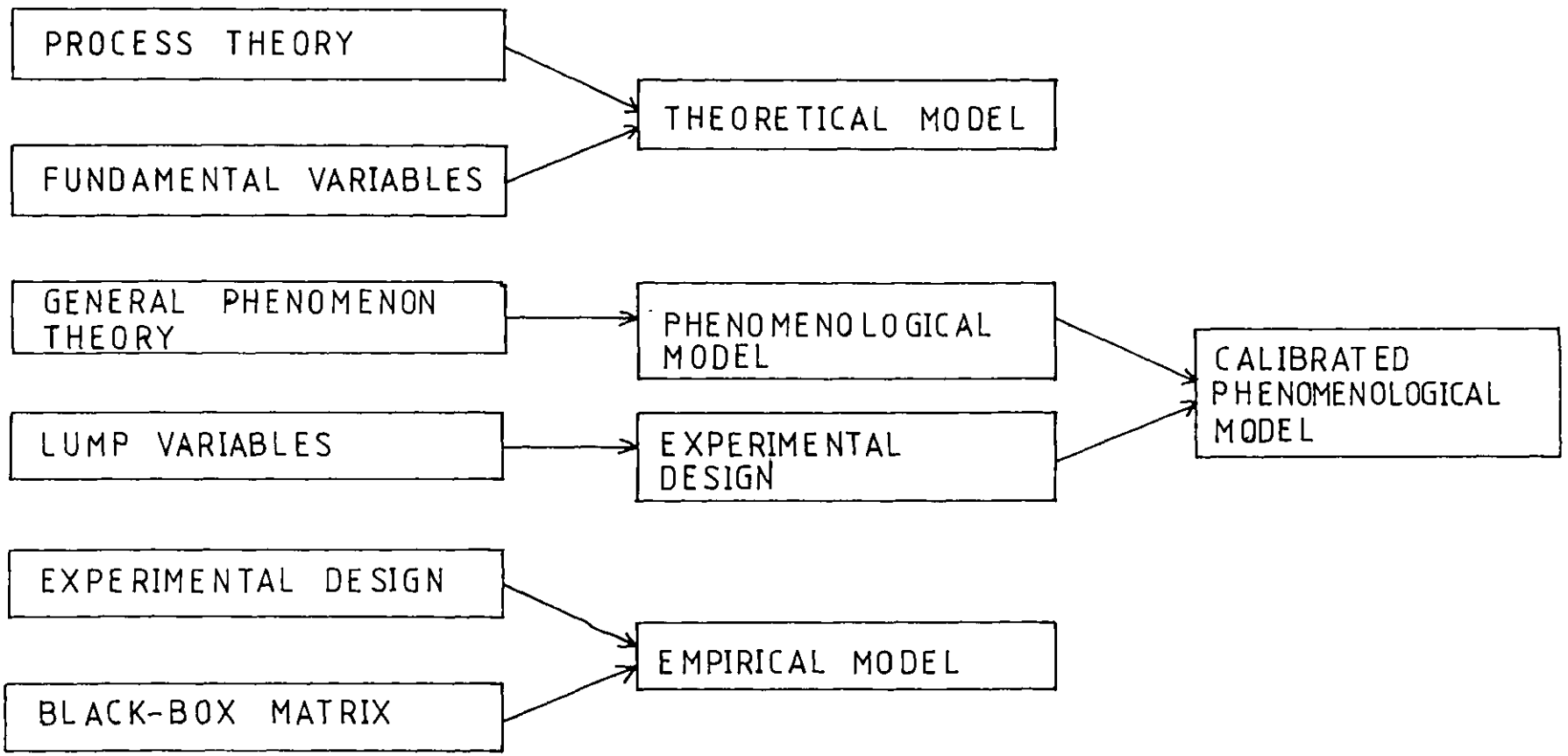


Figure 2.1 The 3 main families of process models

also, for some, a major effort needs to be put into establishing accurate estimation of parameters. Another limitation is the character of the tools available to manipulate the mathematical statements that compose the model (12). However, perhaps the most important problem is the realization of the restrictions of the technique by its users; the models and simulations are often endowed with validity and accuracy above that which they merit (38) and are extrapolated dangerously beyond the range of the original variables (12).

2.0.2 Stream Characterization

The nature of unit operation models in most process plant simulation systems is to calculate the output of an operation given a description of the input stream variables and the process operating variables; a model of a complete process is built up of a number of unit operation models linked together by a number of streams, with the streams carrying the information describing component flows and properties. Therefore before studying the modelling of individual unit operation processes, it is necessary to consider the stream data structures and how component flows may be characterized; it should be noted though that the issue of stream data description is also related to program structure and calculation procedures, the discussion of which is deferred until Chapter 4.

In chemical process simulation it is common practice for only component flow rates and overall stream properties, such as temperature and pressure, to be handled in the stream structure and for other component properties to be stored and/or calculated through global data banks and standard property estimation routines. Whilst the packages associated with a chemical plant simulator would be largely redundant in a simulation of a minerals extraction process, this general method for handling stream data would still appear to be appropriate although new property routines would need to be developed.

When describing liquid-solid slurries, a logical distinction to be made is between component flows in the liquid and solid phases; whilst certain stream properties are common to both, the description and calculation of other important properties will be quite different. The mass flow of components in the liquid stream can be represented in

terms of either the total mass flow and the concentrations of each chemical species, or the mass flow of each species. The properties of the liquid phase can then be determined from the component mass flows and overall stream properties.

Solid component flowrates and properties can also be calculated in this fashion, but such a procedure would be tantamount to assuming a totally homogeneous particle population; it is also necessary to describe at least the mineral content and size of each particle. Other particle properties such as surface area, shape, specific gravity, magnetic susceptibility and conductivity also need to be defined to describe a particle completely; however, whilst these properties can vary between particles of the same size and grade, it is often assumed that this variation is negligible and that these two primary properties allow any secondary property to be estimated with reasonable reliability. As will be seen in the modelling of flotation, some workers have found it necessary to introduce a third property parameter to allow for variations in results; this is a relatively ignored field in modelling, though, and is one which should merit further investigation, especially as and when phenomenological models of processes such as magnetic and specific gravity separations are introduced. For the present though it shall be assumed that if particle size and grade are defined fairly rigorously, then the variation of properties for a particle of a particular size and grade will be negligible. It is felt that this is a reasonable assumption giving considerable flexibility whilst maintaining economy of computer storage space and parameter definition.

Having decided on the properties, size and grade, to be defined, it is necessary to formulate a method of describing the distribution of these properties within a particle population; it is immediately obvious that it is totally impracticable to describe each particle individually and that therefore particles of similar properties falling between certain limits must be grouped together. This would be a discretized distribution, though if the limits are infinitely narrow then the distribution will be continuous.

Continuous distributions of particle size are quite commonly used in the mineral industry, but no such distribution has been found to describe any particle population over the entire size range with

sufficient reliability. This problem is accentuated in process simulation when the recycling and combining of streams are likely to cause distortions to and deviations from the standard functional form. Similar distributions, often called liberation models, have also been proposed for the description of particle mineral content. However models such as Wiegel (42, 43) and King (44, 45) are not only mathematically incorrect and restricted by the assumptions made in their development (46), but they also only define the degree of liberation in terms of the proportion of particles which are completely liberated; this is not particularly useful, especially as it provides no information on the distribution of mineral content values throughout a particulate assembly. Some authors (47 - 49) have discussed the use of graphical quantities to represent an effective degree of liberation, defined by mathematical forms used to approximate grade recovery or release analysis curves obtained from a laboratory separation process such as heavy liquid separation or flotation. However, these curves also reflect other properties which control the separation process and rely heavily on its efficiency, and therefore their interpretation requires a detailed knowledge of the behaviour of the relevant separation process; in the absence of such knowledge they provide rather ambiguous measures of the liberation distribution. Also they are only appropriate to the particular particle size distribution used in the laboratory tests. The only valid description of mineral content of general and flexible enough form to be suitable for process simulation was proposed by Andrews and Mika (50); however the model parameters are difficult to characterize and measure, and the lack of an appropriate analytical solution of the model equations makes it too unwieldy for practical use in simulation (50).

Due to inflexibility and inappropriateness of the proposed continuous distributions it has become customary for unit operation models in the minerals industry to be formulated around discretized size-mineral grade distributions. Using such a distribution it should be possible to derive particle class properties such as specific gravity, magnetic susceptibility and, probably, floatability; the more narrowly defined the particle classes are, the more accurate a simulation based on them should be.

Having determined the structure of the stream data structure it is still necessary to provide values to describe the feed stream of the process being simulated. Component mass flows and overall stream properties should be relatively easy to define, as should the particle size distribution which can be measured using screen and possibly cyclosizer analyses. However the definition of the mineral content distribution of a particle size class is likely to prove considerably more arduous; the two types of analysis technique are laboratory simulation of a separation process and particle analysis using optical or similar techniques. The former depends very largely on the efficiency of the process for the material under consideration, although heavy liquid analysis very frequently proves effective; its main deficiencies are when studying fine particles (<50 microns) and if different mineral phases have very similar densities. Optical and similar techniques are quite commonly used although they can be quite time consuming and expensive, and suffer from the inaccuracies of transforming data measured in one dimension to relate to three dimensional properties; Jones (51) has recently reviewed these methods, both traditional and modern, and highlighted their deficiencies and problems.

2.0.3 Summary and Conclusions

In steady-state process simulation the role of the unit operation model is to calculate the outlet stream properties from descriptions of the input stream variables and unit operation operating variables; the only parts of a process which need to be modelled are those which have an effect on the process mass and energy balance, thus excluding equipment, such as conveyors and bins, which acts only in a material handling role.

Acceptance of the mathematical modelling of processes has only slowly been accepted within the minerals industry, due to some conservatism in ideas, but largely because of the complexity of the minerals and the processes used to extract them. Whilst this situation is improving, the development of useful theoretically based models to predict process performance from known mineral parameters is something of the future, the most suitable models being those which approximately represent the most important properties and which rely on testwork on the material being studied to be carried out to supply values for the model parameters. There are in use an increasing

number of phenomenological models which combine the theoretical and empirical approaches; these describe the contribution of the important process variables to the overall performance and contain adjustable coefficients which must be calibrated against experimental data collected under appropriate conditions. This type of model is preferable to totally empirical models, which are generally only useful over a limited operating range and offer little insight into the mechanisms of the process; however in certain instances it is likely that for a process design application empirical models will be preferred to phenomenological models as the necessary testwork to define parameters for the latter may be excessively time consuming and/or require large scale experimentation of an unsuitable kind. Due to the effort necessary to develop and test models to a suitable standard, it was found advisable to use available published models rather than to attempt to generate new ones.

Whilst perhaps not directly affecting the basic unit operation model structure, the characterization of the stream components is important as it is the process streams which are used in a simulation to carry information between process units. The description of fluid streams is comparatively straightforward as most properties can be deduced from the mass flow rate of each component, overall stream conditions such as temperature and pressure, and characteristic component properties which may be contained in a data bank or, alternatively, as a part of the stream information.

However streams containing minerals are more difficult to characterize due to the heterogeneity of most particle populations. Whilst it could eventually prove desirable to include a variety of solids' properties such as surface area and shape, it would appear that the most useful solid stream data structure, in terms of flexibility and simplicity, should describe particle size distribution and mineral content distribution. Due to the impracticality of describing every individual particle and the lack of suitable continuous distribution functions for size or mineral content, it is best to formulate unit operation models around discretized distributions; if the constituent particle classes in such a distribution are narrowly defined, then the variation in properties between particles of a class should be negligible.

2.1 Comminution

Comminution is a process whereby particulate materials are reduced from various feed sizes by crushing and grinding to the product sizes required for downstream processing or end use (52). It has been estimated that of all the energy artificially generated in the world, about one twentieth is consumed in the size reduction of solids; in the primary metalliferrous industry alone, nearly 2500×10^6 tonnes of ore were treated annually in the early 1970's (53).

The purposes for which solids are comminuted are (54):-

* to decrease particle size
* to increase material surface area
* to free material from its matrices.

Although there are numerous means of achieving these purposes, the minerals extraction industry normally classifies the available equipment into crushing and grinding equipment; it should be remembered however that primary ore breakage by explosive shattering or mechanical excavation is also of importance in the comminution process.

2.1.1 Preliminary Considerations in Comminution System Design

Size reduction is invariably a cost intensive operation, usually the most expensive in a minerals extraction operation, and so careful preliminary consideration to overall system design is always economically worthwhile.

Initially it is necessary to consider the nature of the required comminution products, especially the number and particle sizes. In metallic ores this is dependent on the liberation size of the valuable minerals and the nature of the extraction process; in some ores it may be possible to pre-concentrate at coarse sizes, in many processes it is necessary to almost totally liberate the valuable mineral to produce high grade products, whilst in others it is only necessary to ensure adequate exposure of a mineral at particle surfaces. When treating many coals and industrial minerals it is only necessary to achieve relatively limited size reduction coupled with sizing to give a number of marketable products.

There are two main considerations in the design of comminution systems, the selection of the type and the selection of the capacity of the machinery. Whilst the latter depends largely on the inter-

pretation of testwork results, the selection of the flowsheets that would be suitable for a particular ore in a particular situation is still almost totally reliant on the designer's experience and non-numerical techniques. There are a number of factors which should be evaluated to provide background information for design purposes and to help decide whether a comminution flowsheet will be operable in an efficient manner; apart from macroscopic site and operating factors, such as the nature of run-of-mine ore, the range of tonnages per shift and the number of operating shifts in a week, characteristic properties of the ore which should be examined include abrasiveness, toughness and stickiness, microscopic and mineralogical structures, and physical properties which may indicate how the rock will break.

Once the initial study of the deposit has been completed, it should be possible to narrow down the choice of suitable comminution routes and in the metals extraction industry this will include gyratory, jaw and/or cone crushers and ball, rod and/or autogenous mills; for softer non-metallic ores, impacting machines such as hammermills are often used.

There will still be a number of important design decisions to be made, such as on machinery size and throughput, flowsheet arrangements, process economics and the interaction of comminution processes with the extraction processes, and for these it is necessary to carry out testwork on the ore and more detailed evaluations; a number of authors have described appropriate approaches to these evaluations and the methods involved in determining ore characteristics (54 - 59) and are discussed in more detail in the appropriate sub-sections of this chapter.

Even with this restricted range of equipment to be considered, selection of the best circuit to crush and grind minerals to the size analysis required is a complex problem. Circuits which were considered conventional one or two decades ago are not now suitable under present day economic conditions (60); for instance it has long been held that crushing is cheaper than grinding over appropriate size ranges (e.g. 59), but now, with increases in mill sizes and power utilization, the veracity of this maxim is more doubtful (58). Selection of the

comminution system capable of achieving the best products at the lowest cost for a specific ore and plant is increasingly being sought by improved equipment and circuit design techniques and better testwork and computer analysis methods (60). However, despite the enormous amount of research in the field of comminution (Bickle (61) gives over 2,800 references and the IChE bibliography (62) contains an additional 4,000), the processes involved are still only partially understood, and design procedures tend to be largely empirical.

Despite the differences in the actions of the various comminution machines, there are some similarities in the mathematical models often used for predicting energy-size reduction relationships and product size distributions, though these are still largely empirical and can at best be described as phenomenological.

2.1.2 Energy - Size Reduction Relationships

Probably due to economic considerations, much of the earlier research into comminution was concerned with the relationship between energy consumed and the amount of size reduction that the consumption of this energy brought about (53). Various workers have observed that, in a size - reduction process, the small size change produced was proportional to the energy expended per unit weight of particles, and that the energy required to bring about the same relative size change was inversely proportional to some function of the initial particle size. The relationship between energy and breakage may be expressed as:-

$$dE = - C \frac{dx}{x^{\alpha}}$$

where E = energy consumption per unit mass

x = particle diameter

α, C = constants

The constant 'C' is generally held to be specific to the ore under consideration; however the value of the exponent 'α' has been the source of a great deal of controversy and has been subject to various interpretations. In the 19th century Rittinger suggested that energy consumption should be proportional to the new surface area produced, whilst Kick postulated that for the same degree of fracture in geometrically similar bodies, the energy required is proportional to their volumes; it is now common practice to regard these 'laws' by

substituting 1.0 and 2.0 for 'α', respectively, and integrating the differential equation. A third 'law' of comminution has been proposed by Bond (63), based on an analysis of a considerable amount of data from the Allis Chalmers company; Bond postulated that energy input is proportional to the new crack length produced, and that there is a continual accumulation over successive comminution stages of the energy available for further reduction in the form of incompletely exploited cracks. This 'law' can be represented by setting the exponent 'α' to 1.5. Some workers have tried to show that each of these theories are correct but only over a limited size range, as in:-

$$dE = - C \frac{dx}{x^{\alpha} f(x)}$$

Even this has raised arguments, Hukki (64) proposing one scheme whilst Rose (65) drew an opposite conclusion. The variation in the value of 'α' has been recognized by Charles (66) who allows 'α' to vary between 1.5 and 2.5, which should therefore provide a better description of a given system. However Bond's theory has been supported by considerably greater investment, especially in the collection of data for appropriate scaling factors, etc., and so it is more widely used in mill design. With the use of various factors Bond's equation has been found to be a useful predictor of power requirements, particularly in the sieve size range, and with the use of standard test procedures is of great practical value in mill design.

As more experience is amassed and with the growing availability of reliable analyses, a more cautious attitude to the use of energy-size reduction relationships is now developing (55); this is especially warranted by the errors and shortcomings in the design of some comminution systems based on this approach (67, 68). With the waywardness of energy in a comminution machine, into sound, heat, friction and deformation as well as breakage, it is somewhat remarkable that these theories bear any relationship to the energy consumed in size reduction. Any claims to representing the actual comminution process should be ignored and the equations, especially Bond's, used as purely empirical correlations of operating data, albeit useful equations. Other drawbacks of these equations are that they fail to account for any of the important subprocesses - breakage kinetics, material transport and size classification - in an explicit manner and that they only predict a single measure of the product size

distribution, implicitly assuming a particular form for that distribution.

2.1.3 Particle Size Distribution Models

Due to the limitations of the energy-size reduction relationships there has been a great deal of research effort expended recently on the development of phenomenological models to predict the size distribution of the product from a specific feed during a comminution operation. These models have largely been based on a description of comminution processes usually attributed to Epstein (69), though also similar to Kolmegerov (70). They involve the use of two functions which were termed the selection function (S) and the breakage function (B) in a formulation proposed by Broadbent and Callcott (71, 72). These represent the probability of breakage for a given size of particle in a given event and the resultant fragment size distribution from that breakage event, respectively. The size distribution of a particulate population is normally divided into a number of finite, narrow size intervals which allows the models to be formulated using matrix notation as:-

$$P = (BS + I - S)^{m}F$$

where P = mass of particles after m breakage events in each size interval

F = mass of particles before breakage in each size interval

I = Identity matrix

The models have mostly evolved from probability concepts into a kinetic form where the three main comminution mechanisms are (14):-

R = rate at which particles are broken

A = manner in which the fragments of breakage appear in other size ranges

D = rate at which particles are discharged from the breakage machine

and

$$P = D.R^{-1}(I - A)^{-1}(F - P)$$

The appearance of function A is almost entirely related to the ore properties and the breakage rate function R is primarily related to the energy input per tonne which is machine dependent as well as being related to the rock properties.

2.1.4 Liberation

One of the major reasons for comminution is to liberate the valuable mineralogical species from those which are not valuable. However, most research has been on the comminution of particle assemblages as opposed to characterization of mineral release during comminution processes; most modelling effort has been directed towards predicting particle size distribution and energy requirements (73). To date, the manner in which minerals are released from the host rock has proven resistant to mathematical modelling techniques; as has been shown, even the mere characterization of the degree of liberation is difficult to either define or measure accurately. As Finch et al (74) have highlighted, whilst batch laboratory tests can be used to simulate and help predict changes in overall particle size distribution in industrial scale operations, there are significant quantitative scale-up differences in respect of mineral release and the effects of selective grinding. The latter phenomenon occurs as soft material is reduced considerably more quickly than the harder material in a particle population and so as the comminution process proceeds the hard particles remain large longer, thereby shielding the softer material.

In their analysis of the comminution of a heterogeneous material Andrews and Mika (50) used a continuously distributed system of size and grade; however the models developed are of limited value as, due to their complexity, it is not possible to find solutions to them which possess general validity. Also the distributions used are difficult to characterize and become excessively unwieldy in simulation. King's model (44, 45) has been used by Finlayson and Hulbert (75) to simulate the behaviour of individual minerals in a closed grinding circuit; selection and breakage functions are determined for pure minerals and the functions for locked particles are estimated by linear interpolation. However, as has been shown (46), King's model is fallacious and can only describe the fraction of material unliberated. A similar interpolation approach was used by Cutting (73) to predict the work indices of quartz-magnesite mixtures. Whilst this type of approach reduces the task of determining parameters for each category or grade of mixture, it is one which is likely to be fraught with possible problems. As has been seen, selective comminution occurs and so changes the breakage character-

istics during the comminution operation, a situation which is likely to be difficult to describe numerically in steady state simulation models. Also breakage of heterogeneous particles is likely to occur along grain boundaries rather than across them and interpolation of the breakage properties of homogeneous particles will not account for this.

Whilst predictions of mineral release in comminution using results of laboratory scale testwork are inaccurate, it is most unlikely that models of this process will be developed. However it is necessary, due to the importance of liberation in subsequent processing steps, that such models are developed, perhaps using data from larger scale operations. It should be remembered though that, not only are liberation characteristics likely to be unique to the material being studied, but also that, due to the different mechanisms of breakage, they will also be specific to the type of machine used. An approach which might produce results would be to investigate how heterogeneous particles break relative to the breakage properties of their component minerals; this could lead to procedures for interpolating between the characteristic properties of the pure components However, of prime importance, it is necessary to develop testwork methods which can reproduce industrial scale processes in the laboratory.

Despite these problems it is still necessary to describe the output streams of a comminution process for simulation purposes, as it is also necessary in conventional metallurgical design procedures to produce feed material for testwork to help design subsequent treatment processes. It would appear that it would be best to treat the material in the comminution process as a homogeneous particle population and to define the effects of liberation subsequently, based on the results of testwork and some experienced judgement. As has been seen no practical functions can be used to describe mineral distributions and so it would be necessary to define the proportion of material in each discrete size and grade class; a suitable method might be to define the fraction of material in each size class, which is in each grade class and to allow the comminution model to predict the fraction of the total particle population which lies in that size class.

2.1.5 Crusher Design and Modelling

Despite the complex engineering problems and high costs involved in the design of crushing facilities, this always has been and still remains one of the most technologically neglected areas in mineral processing (76). This is often because crushing is regarded as a poor relation to, and only as feed preparation for, the grinding circuits, where tremendous efforts have been expended to improve understanding. However the crushing circuits often involve the highest capital expenditure in a treatment plant and large cost savings can be made both here and in the grinding circuits with their optimum design and use.

Although numerous crushing machines exist and are in common employment, when considering metalliferrous mining two types predominate, that is the jaw crusher and the gyratory crusher; cone crushers are used almost exclusively for secondary crushing and are very similar in design to the gyratories. These machines have been in extensive use for many years and their description and discussion would only repeat many previous works (e.g. 52, 54, 55, 77, 78). Although, as is perhaps to be expected, manufacturers often claim tremendous advantages for their own products, crushers are generally very similar and in preliminary analysis can normally be regarded as identical; however, this is not to say that when specifying and purchasing plant for a particular installation that a certain machine does not have any advantage, it is only a reasonable approximation that this is not so.

The design of new crushing facilities is largely based on the analysis of accumulated equipment performance characteristics and the results of empirically based and relatively crude testwork. Bond's work index is normally used to indicate energy requirements or crusher product size. For coarse crushing the Bond Crushability test (55) is used to give a measure of the impact strength of a material; however these tests are difficult to replicate and can indicate different work indices for different size fractions (79). As this impact work index customarily gives bad correlation for fine crushing (79), the rod mill work index (55) is used in preference. Whilst it may be considered logical for crushing testwork to be carried out on run-of-mine ore (58), especially considering the arguments about the use of the energy-size reduction relationships over various size ranges (55, 64, 65),

this is rather impractical and unnecessary as smaller, competent rocks, upon examination and under test conditions can show similar characteristics (58). Most equipment manufacturers are equipped and willing to carry out the necessary testwork and pilot plants are normally only used for potentially difficult ores.

Crushing tables as supplied by manufacturers (80 - 82) are normally used to size crushing equipment for a given flowsheet; these give capacities and product analyses for a particular machine using relationships established within the industry, which, whilst empirical, are accurate enough for most design purposes. Although well tried and trusted, these methods are still prone to error, such mistakes being difficult and expensive to correct (e.g. 83), and are best used by experienced design staff.

A major research effort on crushing plant design has recently been carried out by the Allis Chalmers company, with some synopsis of the work by Flavel in a series of papers (76, 84-86) and a full description of the principles and mathematics involved given in a comprehensive paper (79). Whilst claiming to provide a completely new fundamental approach to crushing plant design, the proposed method is still based on Bond's equation and uses empirical product size distribution relationships. However, rather than using crushing tables, Flavel introduces the concept of the "power rate of a crusher", this parameter being defined as the ability to inject energy per unit of feed and is largely dependent on the crusher chamber configuration and control of the closed side setting. The major assumption made in this approach is that the crusher has an adequate amount of evenly distributed feed, which requires the provision of suitable feed mechanisms, a design feature which is being increasingly appreciated regardless of the procedures used to size the facility.

Whilst some effort will be required in determining the power rate of different crusher arrangements and types, it is likely that this approach will soon begin to supercede the use of crushing tables, because a crusher's power rate is machine specific whilst the capacity tables do not define the feed size and work index to which the capacities relate and being more material property related, are not as accurate

when used to design crushers treating rock different to the reference material.

Once a material's work index is known, the energy required to reduce that material from its feed 80% passing size to the necessary product size can be determined from Bond's equation:-

$$W = K.Wi\left(\frac{1}{\sqrt{P}} - \frac{1}{\sqrt{F}}\right)$$

where W = energy consumption per unit weight

Wi = material work index

P,F = product and feed 80% passing size

K = constant involving mechanical and process efficiencies that will be quantified when a new crushing work index parameter is established; presently taken to be equal to 1 (79).

The average tonnage to the crushing plant is calculated from the average mine production rate and an estimation of the availability of the plant over a period of one year; from this the total crushing plant power rate can be estimated and subsequently, using knowledge of the average power conditions prevailing and crusher power rates, it is possible to calculate the necessary number of crushers. The crusher product size distributions are determined using standard discharge curves as published by Allis Chalmers (80) and of similar form to other manufacturers' curves (81, 82). Alternatively the power rate and number of crushers can be determined from the crushers' installed power and design feed capacity.

For use in a simulation plant model, both Flavel's method and the more traditional procedures require some method of determining the crusher discharge particle size distribution. As has been seen, techniques presently rely on the use of standard discharge curves, which are totally empirical and the basis for which is not given. From the appearance and use of a large family of similar curves it is reasonable to postulate that regression equations could have been used in their production; this has perhaps been confirmed by the publication of such a set of equations by a large manufacturer for gyratory crushers (87). A number of equipment manufacturers were then contacted and details of such equations requested; however no such details were

received, although two companies implied that existing published tables and curves would continue to be used (81, 82). Using the techniques published (87) for gyratory crushers as a model, data from two equipment manufacturers' publications (80, 81) were plotted as in Figs. 2.2 - 2.5 and empirical equations determined using regression analysis and the basic form:-

$$P = a X^b$$

where P = percentage passing the screen size Y

X = (size of square screen opening Y)/(topsize of product T)

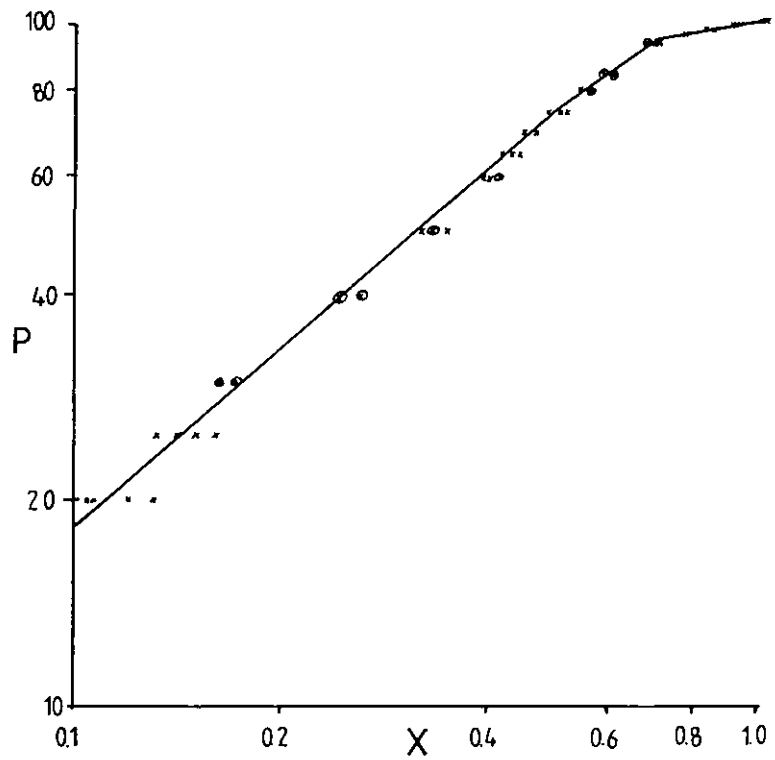
a, b = regression constants.

As can be seen from the graphs, the curves for each manufacturer are virtually identical, thereby perhaps disproving their claims of superiority for their own machines, and so the equations were regressed from the combined data sets; the data was analysed using MINITAB (88) and equation parameters, with relevant statistical data are shown in Figs. 2.2 - 2.5. The major drawback of using these equations, other than the nature of their determination and basis, is the need to use three different equations and so having discontinuities, the positions of which were estimated by inspection and some trial and error; other equation forms were used, as were different discontinuity positions, but those equations described were the best of those tested. Also the overall product size distribution is only indirectly related to the feed through the 80% passing sizes. The maximum size, T, for a particular distribution can be calculated by substituting in the 80% passing size predicted by Bond's equation.

Empirical relationships apparently based on traditional design methods and data have been used in crushing plant simulation programs by a number of authors (89, 91); however few details of the models so derived have been published, as have few examples of their use, and so their general validity and acceptance are difficult to evaluate.

As has been described, the representation of comminution processes in terms of the breakage probabilities and primary fragment size distributions was suggested by Broadbent and Callcott (71, 72). Their formulation was based on the probabilities related to one breakage event and therefore requires some estimation of the number of events likely to occur. Whiten (92) extended this concept to include internal particle classification and hence allow for the residence

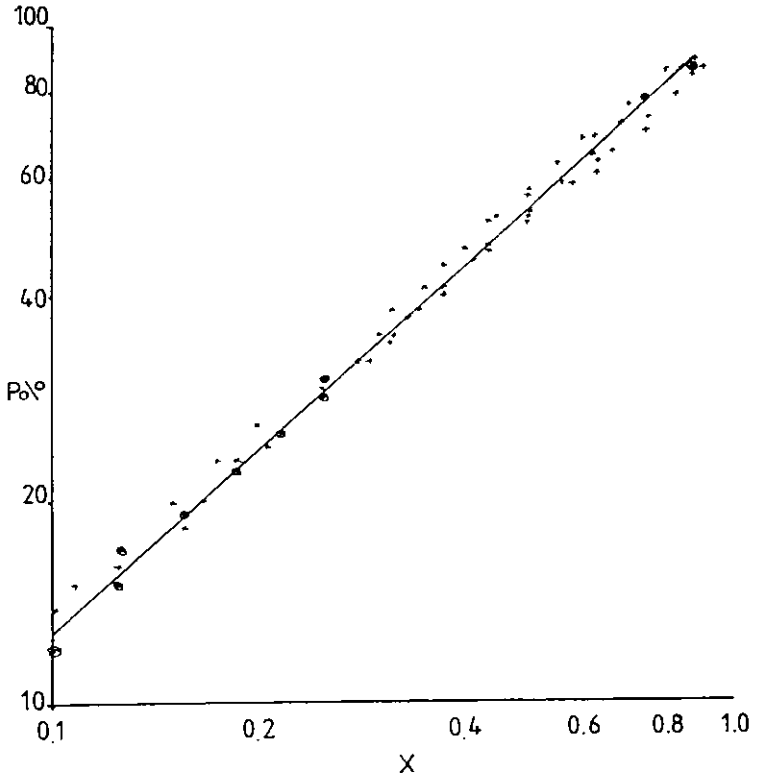
Figure 2.2 Gyratory Crusher Product Size Distributions (87)



$$P = 136(X)^{0.878} \quad \text{when } 0.01 < X \leq 0.5$$

$$P = 121(X)^{0.711} \quad \text{when } 0.5 < X \leq 0.7$$

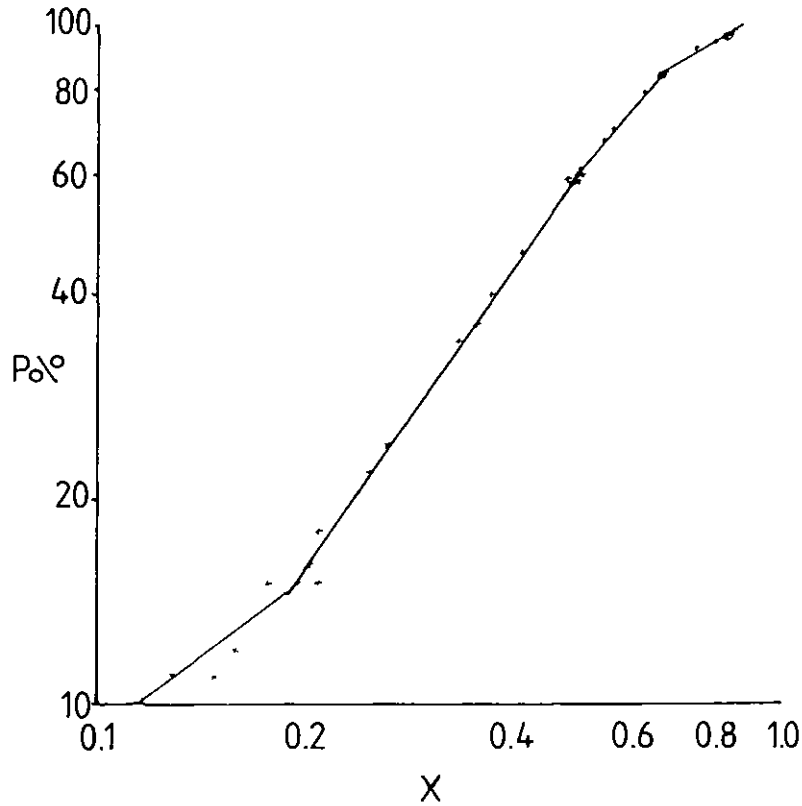
$$P = 100(X)^{0.173} \quad \text{when } 0.7 < X \leq 1.0$$



$$P = 100(X)^{0.902}$$

R=99.2% Std.dev. = 0.0371

Figure 2.3 Jaw Crusher Product Size Distribution



$P = 50.11(X)^{0.741}$ when $X<0.185$

$R^2=95.4\%$ Std.dev. = 0.0338

$P = 158.49(X)^{1.41}$ when $0.185 \leq X<0.5$

$R^2=99.1\%$ Std.dev. = 0.04

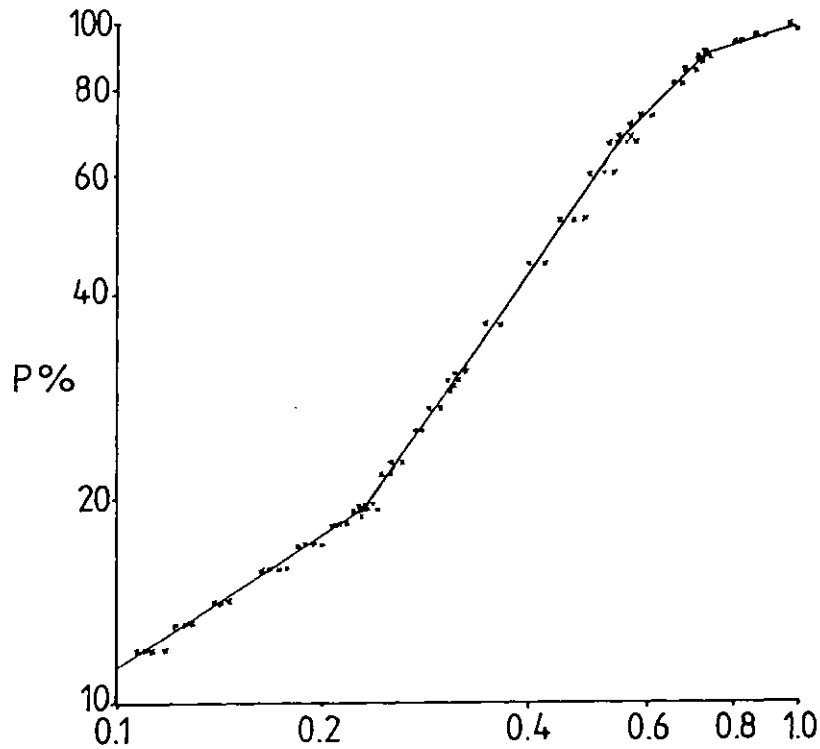
$P = 131.83(X)^{1.12}$ when $0.5 \leq X<0.67$

$R^2=99.3\%$ Std.dev. = 0.007

$P = 107.15(X)^{0.6}$ when $0.67 \leq X$

$R = 98.3\%$ Std.dev. = 0.004

Figure 2.4 Closed-circuit cone crusher product size distribution



$P = 77.6(X)^{0.939}$ when $X < 0.23$

$R^2 = 99.0\%$ Std.dev. = 0.008

$P = 162.2(X)^{1.43}$ when $0.23 \leq X < 0.55$

$R^2 = 99.9\%$ Std.dev. = 0.004

$P = 123.0(X)^{0.951}$ when $0.55 \leq X < 0.72$

$R^2 = 100.0\%$ Std.dev. = 0.0

$P = 102.3(X)^{0.377}$ when $0.72 < X$

$R^2 = 98.0\%$ Std.dev. = 0.002

Figure 2.5 Open circuit cone crusher product size distribution

time within the actively crushing part of the machine. This was by means of an arbitrarily functional form based on Gaudin's probability of a particle passing a screen (93) and related to variables such as closed side setting, fraction of material larger than one inch and feed tonnage; also particles larger than an arbitrary size (K_2) are always broken to a finer size and particles finer than another arbitrary size (K_1) pass through the crusher unbroken. Whiten's model is of the form:-

$$P = (I - C)\,(I - B'C)^{-1}F$$

where I = Identity matrix

P,F = product and feed flowrates

C = internal classification matrix

B' = overall breakage matrix

and C_{ij} = $\dfrac{\text{tph of size i entering product}}{\text{tons of crusher content of size i}}$

$$C_{ij} = \frac{\int_{d_{j-1}}^{d_j} (1 - \{(d - K_2)/(K_1 - K_2)\}^2)\,d(d)}{d_j - d_{j+1}} \qquad K_1 < d < K_2$$

$$B_{ij} = \frac{\text{tph entering size i from size j}}{\text{tph of size j not leaving product stream}}$$

$$= (I - (I - B)P)$$

$$B = \alpha B_1 + (1 - \alpha)B_2$$

$$B_{1ij} = (1 - \exp\,(d_i/d_j)^6)/(1 - \exp(-1))$$

= cumulative fraction finer than size d_i resulting from impact breakage of sized j

$$B_{2ij} = (1 - \exp(d_i/K)^r)$$

= cumulative fraction finer than size d_i resulting from abrasion breakage of size d_j

K_1 = arbitrary size below which $C_{ij} = 0$

K_2 = arbitrary size above which $C_{ij} = 1$

K_1 and K_2 are dependent on closed side setting, fraction of material larger than one inch and feed tonnage.

α, K and r are empirically determined constants.

This is almost entirely empirical, and in places borders on the arbitrary; however it has been successfully used to simulate crushing circuits (53), although it was found necessary to use new empirical expressions for α, B_1, B_2, K_1, K_2 in another circuit analysis (94). Whilst the basic form of this model has proven useful in simulating existing operating plants, a more general basis is required for determining the breakage matrix values and empirical constants in the internal classification matrix; if this model is to be used in the design of new plants, it would be necessary for these parameters to be determined under laboratory test conditions which at present is impracticable.

Rose and English (95) attempted to model the performance of jaw crushers theoretically based on their dimensions and crushing action; however empirical assumptions had to be made with respect to the speed of the crusher face and the energy efficiency and the resulting model appears to have found little favour industrially.

A phenomenological approach to crushing characteristics, similar to Broadbent and Callcott but including energy considerations, is being used by Krogh (96) to develop a model whose parameters may be estimated from tests carried out on single particles. From such tests, three fundamental crushing functions may be determined; these are:-

1) Probability Function: the probability of crushing a particle at a certain energy input.
2) Energy Function: the energy necessary to break 50% of the particles as a function of particle size.
3) Breakage function: size distribution of fragments from the crushed particles when energy input corresponds to a crushing probability of 0.5.

By allowing for the dependence of particle breakage on energy considerations and taking into account the need for model parameters to be measured in small scale tests, it is feasible that this approach may develop a useful design tool and a means for studying more closely the crushing mechanism. However there is still considerable development to be carried out before this goal is achieved.

2.1.6 Grinding Mill Design and Modelling

Whilst all comminution operations are cost and energy intensive, grinding mills tend to be used at the finer end of the size spectrum and are major consumers of energy and steel as well as requiring major capital investment; indeed these two operating costs tend to be two of the largest expenses in a mineral extraction plant's operating budget. There are three commonly used grinding systems available in mineral processing, ball mills, rod mills and autogenous mills; with these alternative systems available and an array of possible combinations and arrangements, the task of designing a grinding plant capable of producing a satisfactory product at minimum capital and operating costs is complex and difficult. Traditionally, design methods have been empirical and based on specific energy concepts such as Bond's equation, although the use of more rigorously based techniques is becoming increasingly widely accepted.

As has been stated already, in the metals extraction industry comminution is invariably carried out to release minerals from the constituent rock matrix or to expose mineral surface for chemical access. Thus the grinding requirements are governed to a large extent by liberation size and the design criterion is to reduce a tonnage of ore to that size economically. However it is often more expensive to release and recover an increment more metal than its market worth and so evaluations should be carried out at different product sizes to determine the optimum degree of grinding; such studies and methods for carrying them out have been suggested by a number of authors (97 - 99). Simple functions relating recovery to grind size and grinding costs to specific energy are developed and form the basis for calculations; not only are these functions approximate, but their basis is debatable, and techniques which allow more detailed analysis of grinding variables and costs with recovery variables, costs and revenues would therefore be particularly useful.

2.1.6.1 Ball and Rod Mills

Current design procedures are based largely on empiricism using correlations of mill performance with some kind of grindability test; the most common tests have been described and discussed by Marshall (55). Whilst a preliminary indication of grinding characteristics can often be given based on cursory inspection of the material and simple grinding

tests, more precise assessments require grindability studies and comparison tests made in standard, laboratory sized mills. The most common tests are based on standard procedures to determine an index of grindability, which is then used to select the appropriately sized mill for the desired feed and product specifications and production rate using correlations based on accumulations of empirical data; the most commonly known design methods of this type are Bond's and the Hardgrove method.

Rowland et al (100 -102) have provided details of calculation methods for sizing conventional ball and rod mills based on values of the Bond Work Index. Efficiency factors based on mill variables are used to correct the work index, which is a measure of grinding energy per unit weight, and the size of mill required for the design tonnage is obtained using the corrected work index. As with Flavel's study of crusher design using Bond's methods (79), Allis Chalmers sponsored a major study to compare laboratory determined Bond work indices with work indices obtained under operating plant conditions (100); good correlations were obtained although some of the design guidelines were slightly modified subsequently (102). Unfortunately there has not been any industrial sponsorship and advertising of similar studies on the use of other design techniques of this type, and so their performance is not as commonly known or understood.

Whilst these design techniques allow reasonable confidence in the prediction of performance and operating requirements and are very well established as standard design tools, it would appear that their use and reliability is being increasingly questioned and more rigorous techniques based on relatively detailed models of the grinding process are being studied and in some cases used for the design of new plants. As with almost any method dependent on parameters determined from testwork, the use of this type of design technique can, and has, resulted in poor designs (e.g. 67, 68); these errors, and the techniques' lumping together of all sub-processes, therefore ignoring mass transport considerations, have been seized upon to support the case for a more rigorous mathematically based design procedure using particle size distribution models of the type described in Section 2.1.3. (e.g. 13, 53, 104, 105). However these new procedures have yet to gain widespread acceptance, although this is probably because of the lack of an adequate data base and experience of their use, and a still

incomplete knowledge of the appropriate scale-up criteria (52); indeed a recent survey (60) of the methods used and recommended by major mill manufacturers showed that, of those questioned, only one company used such techniques, the majority using correlations based on standard laboratory tests and indices. The criticisms of the more empirical methods are valid though, and it is reasonable to assume that they will be increasingly supplemented and eventually replaced by modelling techniques, unless commerical interests hinder their development and use.

It can be shown (106) using conventional statistical treatment that if a random probability of selection for breakage is applied to a large number of particles, the rate of breakage of any particle size is proportional to the amount of particles of that size in the mill; in this way Broadbent and Callcott's probability approach to modelling comminution can be extended to provide a kinetic model of grinding. Whilst there is no a priori reason why grinding should be a first-order process (52), this assumption has been shown to be valid for a large number of experimental ball mill systems (52, 107, 108); however this phenomenon is not exhibited in rod mills and so the following discussion only applies to ball mills.

For a first-order rate comminution process:-

$$r_j \propto W_j$$

where r_j = rate of breakage of particles in size interval j

W_j = mass of particles of size j in the mill

If S_j is taken as the proportionality constant, known as the breakage rate of size j, and b_{ij} is defined as the mass fraction of the fragments produced by breaking size j which falls into size class i, the solution of the differential equation describing the batch grinding process as a function of time is:-

$$m(t) = e^{(B - I)St} m(o)$$

where m = size distribution vector

B = b_{ij} matrix

If the product size distribution from a continuous mill is taken as the average response of a number of batch grinding processes, weighted with respect to the distribution of residence times of material in the mill then:-

$$M_{MP} = \int_0^{\infty} M_{BATCH}(t)\, E(t)\, dt$$

where M_{MP} = product size distribution of a continuous mill

$E(t)$ = residence time distribution of solid material

Therefore:-

$$M_{MP} = M_F \int_0^{\infty} e^{(B - I)St} E(t)\, dt$$

This has been simplified (108) using transformation techniques to give:-

$$M_{MP} = T\, J_c(\tau) T^{-1} M_{MF}$$

where

$$T_{ij} \begin{cases} 0 & i<j \\ 1 & i=j \\ \sum_{k=j}^{i-1} \frac{b_{ik} S_k}{S_i - S_j} T_{kj} & i>j \end{cases}$$

$$J_{cij}(\tau) = \begin{cases} \int_0^{\infty} \exp\{(- S_i \tau)\Theta\} E(\Theta)\, d(\Theta) & i=j \\ 0 & i \neq j \end{cases}$$

$$\Theta = \frac{t}{\tau}$$

τ = mean residence time of mill

$$= \frac{\text{mass hold-up in mill}}{\text{feed mass flow-rate}}$$

It is therefore possible to describe the grinding process in a continuously operating ball mill in terms of three parameters:- the rate of breakage (S_i), the primary breakage distribution (b_{ij}) and the dimensionless residence time distribution of particles in the mill ($E(\Theta)$).

The determination of values of S_i and b_{ij} can be achieved in a series of laboratory tests (108, 109), although the number of experiments can be reduced using a back calculation technique suggested by Klimpel and Austin (110) if a functional form of S_i and b_{ij} is used. Tests on a number of dissimilar materials were carried out, the forms used were suitable for each and so their use would appear not to be restrictive, especially when the advantage obtained in parameter determination and input are considered. Other functional forms have been suggested (e.g. 71, 106) but no comparison or discussion has been carried out on those available.

It is generally assumed that the breakage distribution is an intrinsic property of the ore and does not depend on mill characteristics; it is therefore accepted that its values can be measured in the laboratory and applied to industrial scale mills with confidence. The breakage rate function is, however, a measurement of the grinding efficiency of a mill for a particular material and is, therefore, also dependent on mill characteristics. Despite considerable effort at quantifying this dependence on mill design and operating variables (104, 108, 111, 112), the state of knowledge in scaling-up model parameters is still very limited and so the use of such models in mill design is fairly risky. However for ball mills operating under conditions of dynamic similarity - i.e. with the same fractional ball loading, material hold-up and relative rotational speed - it is found that the specific rate of breakage increases roughly with the square root of mill diameter (52). Herbst and Fuerstenau (109, 113) have shown that rates of breakage are proportional to the specific power input to the mill:-

$$S_i = S_i^E \left\{\frac{P}{H}\right\}$$

where

S_i^E = specific breakage rate function

P = net mill power

H = mass hold-up of material in the mill

The function of S_i^E is independent of mill design and operating variables. Breakage rate functions measured for the same material in mills of different sizes which possess complete dynamic similarity should approximately vary as the root of the mill diameter; therefore if testwork is carried out in a similar mill to that being designed, then scale-up should be relatively straightforward. Another important feature of this result is that it implies that much of the data collected on work indices potentially can be applied to the estimation of specific breakage rates (52) and standard power correlations can be used to account for mill variables. It should be noted though that results have only been published for scaling laboratory results to pilot plant mills and that considerable effort is required before design of industrial mills becomes more reliable.

The third model parameter, the mill residence time distribution, is also an area of incomplete understanding, and characterization of industrial mills is needed. Many mathematical models have been proposed to describe the distribution, amongst them the diffusion model (114), perfect mixers in series (115), perfect mixers with a plug flow component (116) and empirical equations (105). As there has been little work on applying these to different mills and, the choice of a model is therefore somewhat arbitrary, it is necessary to select a flexible approach which has the potential of being valid for different situations. Herbst and Fuerstenau (109) suggest that for a particular mill type, if the ratio of mill length to mill diameter is held constant and dynamic similarity is maintained then the dimensionless residence time distribution, $E(\Theta)$, is independent of the size of mill and scale-up of the model parameters is feasible.

The discussion of particle size distribution prediction models above is suitable for systems where the breakage rate and distribution functions are independent of particle size and time; in this situation the kinetic model is linear with constant coefficients. As was implied earlier this is generally a valid approach for modelling ball mills but is inappropriate for rod mills. The important difference between the two types of mill is the predominant effect of internal classification resulting in strongly size-dependent breakage in rod mills. A rod mill could be regarded as a series of ball mills, each in a closed circuit with a classifier, resulting in a stage of grinding model (53); it should however be remembered that the classification effect and residence time distribution will vary as the flow rate and particle size alter. This therefore requires the development of special transport and classification models, but these, and rules for scale-up procedures, are lacking and so, whilst it is possible to model operating mills, the use of phenomenological models in the design and simulation of rod mills is presently impracticable.

2.1.6.2 Autogenous and Semi-Autogenous Mills

A sector of ore milling which has generated much interest, development and debate in recent years has been autogenous grinding, either full or partial (60). Suitable ore for this type of grinding

is competent and has grain boundaries at the desired product size; breakage occurs primarily along the grain boundaries, mostly by abrasion and attrition, with less impact grinding than in other mill types. Initial testing therefore concentrates on the general nature of the material, especially whether the coarse particles are competent enough to survive as grinding media for enough time. However it is generally agreed that if autogenous grinding is to be seriously considered for a flowsheet, pilot plant testing is required and that techniques suitable for rod and ball mill design are inappropriate; even if such testing is carried out, the methods of scaling-up results to design an industrial installation are very uncertain and serious errors have been made (117, 118).

Similarly, the choice of an appropriate form of model for autogenous grinding is also uncertain, with internal classification and self-breakage of the media being important complicating factors. The effects of some operating variables such as viscosity of the pulp, feed size changes, and the addition of steel balls are not yet accurately known (14), causing difficulties in design and especially in modelling. However Lynch (53) has proposed and used a model for simulating autogenous mills using matrices to describe the appearance of particles in a size fraction, breakage rate and discharge rate and also allowing for mill load. However Manlapig et al (119) have shown that breakage mechanisms may be considerably more complex than in normal grinding, with coarse and fine particles having different breakage mechanisms and that any model of autogenous grinding should account for this phenomena.

2.1.7 Summary and Conclusions

Comminution systems are vitally important elements of virtually all mineral processing plants both in terms of their associated costs and their effects on other plant elements, especially those recovering the valuable components. It is therefore necessary to accurately design comminution systems such that their contribution to process economics and their operating performance can be accurately predicted.

Apart from autogenous mills, comminution machines can generally be designed with confidence, using results of laboratory scale testwork

though it is also necessary to utilize the accumulation of process data made by equipment manufacturers. Whilst the performance of such devices can often be predicted reliably, the selection of a suitable flowsheet combining size reduction and size separation machines and interfacing with minerals extraction processes is a complex task which appears to have little sound basis, depending largely on an engineer's experience and which would benefit greatly from a computer aided approach providing both technological and economic information.

Laboratory simulation of crushing machines is difficult and unreliable, and their design is usually based on the results of rod mill tests. A suitable form has been proposed (92) for modelling the crushing process but some of its parameters use debatable empirical relationships and can presently only be estimated from testwork on operating crushers. Whilst empirical in form and with an imperfect history in design, it is best to base models intended for design on energy-size reduction relationships of the form proposed by Bond, and capacity and product size distribution relationships published by equipment manufacturers. A new approach to this type of design procedure has been proposed by Allis Chalmers, based on the definition of a crusher's power rate, that is its ability to inject power to a unit mass of feed, and power efficiency, and using Bond's formula to predict the amount of size reduction. At present it is still necessary for an engineer to specify a machine's installed power, power utilization and design tonnage, which is virtually the same as using the traditional capacity relationships; as more machines are characterized in terms of power rates, some of these specifications will become redundant simplifying the designer's task. Empirical equations have been derived from manufacturer's publications to describe the product size distributions.

Traditional design methods based on specific energy concepts such as Bond's equation are still largely used to design rod and ball mills. However such methods have drawbacks in that they are empirical, do not describe the product size distribution and do not account for important subprocesses such as breakage rates and transport through

the mill. Therefore more rigorous mathematical modelling techniques are gaining favour, the most popular form of model having three parameters; breakage rate, breakage distribution and residence time distribution. The amount of work involved in obtaining these parameters is similar to that involved in obtaining work indices and provides more detailed information. However considerable effort is still required to define reliable scale-up procedures and relationships and mill residence time characteristics.

The mathematical form of models describing ball mill grinding is fairly standard and well accepted. The actual grinding process is described using breakage rate and breakage distribution functions; these are implemented using matrix techniques, the elements of the matrices referring to individual, discrete particle size classes. A back calculation technique for determining these functions for a particular mill condition and feed material; whilst the technique is based on continuous forms of the functions, it would appear from testwork that these are not of a restrictive nature and, indeed, their use considerably eases parameter definition in process simulation. Breakage distribution is generally assumed to be independent of mill conditions, but it is necessary for the breakage rate function to be measured in a similar mill to that being designed; however progress is being made on the scale-up of this function from laboratory testwork and it is to be expected that more reliable procedures and correlations will be developed in the near future. Development work is also needed on the form of the third model parameter, the residence time distribution; however a number of potentially suitable models have been proposed and it is only necessary to determine how these may be applied to different sizes and types of mill.

Suitable models for use in designing rod mills from laboratory work are not available and so specific energy techniques must still be used. Similarly, there are no appropriate models for designing autogenous mills, even using pilot plant test results; indeed, the design of such mills is generally unreliable and fairly large safety margins need to be allowed.

With the present state-of-the-art all comminution design methods treat the material to be broken as being homogeneous; indeed, it is difficult to simulate mineral liberation characteristics using

laboratory testwork. A great deal of effort and development is required in this area before suitable methods and procedures are available for predicting liberation from laboratory results. At present it is necessary for the design engineer to predict the proportion of each size fraction lying in a particular mineral class after comminution.

2.2 Size Separation

Most minerals separation processes rely on adequate liberation of the valuable constituents by comminution and the preparation of a reasonably homogeneously sized feed. Thus the accurate classification of process streams by particle size, whilst not directly important in terms of mineral recovery and/or cost, can be of crucial importance to the success of a process.

The traditional methods of size classification are based on the age old principle of screening. Most modern crushing plants now use vibrating screens for strictly sizing applications though static screens are useful for approximate, but cheap and robust separations. For smaller particle sizes, screening has largely been surpassed, initially by rake and spiral classifiers and then in turn by the hydrocyclone. Few ore grinding circuits are constructed today without hydrocyclones and, where they are not used, invariably hydrocyclones were considered at the design stage (120). Therefore, whilst static screens and classifiers can be important in certain applications, most systems are based around vibrating screens and hydrocyclones.

2.2.1 Design and Modelling of Vibrating Screens

Size separation using vibrating screens is obtained by repeated presentation of particles to equal sized apertures through 'bouncing' the feed across the screen surface; at each presentation, every particle which is capable of passing through the aperture has a probability of doing so and the higher the number of presentations, the higher is the overall probability that the particle will appear in the screen undersize (53). Descriptions of the types and varieties of suitable machines are to be found in the literature (53, 78, 121).

The selection and design of vibrating screens is invariably carried out based on throughput rate and an analysis of the particle size distribution of the feed stream; even in the detailed design stages, virtually no testwork is carried out as there are few small-scale screens which accurately simulate industrial-scale machines. However as Colman (121) points out, screening is much more of an art than a science, and design procedures, whilst fairly accurate, are based on the accumulation and analysis of large volumes of data by the equipment manufacturers.

There are a variety of design methods, the principles of most having been described by Taggart (78) in 1947 and these have hardly changed since. They all use a series of empirical factors to allow for the effect on separation efficiency of numerous stream variables, related to the feed stream mass flowrate and sieve analysis, and screen dimensions.

Despite this empiricism, there has been a fairly substantial body of literature on the modelling of screen operations although there have been few successful simulations of production plants reported. A number of workers have based their work on developing Gaudin's (93) concept of screening as a probability process. Assuming that the particles are spherical and that no blinding of the screen aperture takes place, the probability, 'p(d)', that a particle of size 'd' will pass through a screen of aperture 'a' and wire diameter 'w' is given by:-

$$p(d) = \frac{(a - d)^2}{(a + w)^2}$$

Canalog and Geiger (91) used an analagous expression, though they assumed the particles to be cubic. This was then related to screen variables using industrial factors from Taggart (78), but the resultant equations are somewhat complicated and must be evaluated numerically.

Whiten (92), in one of the more commonly used models, has extended Gaudin's equation to include the number 'm' of possible presentations of a particle to the screen; the probability, 'p(d)', now describes the likelihood of a particle passing to the oversize stream:-

$$p(d) = (1 - ((a-d)^2/(a+w)^2))^m$$

The exponent 'm' is related to an efficiency constant, screen size and a load factor. In later work, Walter and Whiten (122) found that the efficiency constant varies with the type of screen cloth used and with the volume of feed material. Whilst this is a simple and robust model, it is inadequate in the prediction of sub-mesh material, is not fully tested and requires some experimentation to determine model parameters (15, 53). This is despite the successful use of a slightly modified version by Hatch and Mular (94) in the simulation of a

secondary crushing plant.

Phenomenological approaches describing screening kinetics have recently been proposed by Kapur et al (123) and Ferrara et al (124, 125). Whilst potentially of use in the future, especially in dynamic modelling, their complex structures are not yet suitable for steady-state simulation and further research is required to relate the kinetic constants to screen variables and allow their use in a design context (125).

Another study with similar potential, but also similar problems, has been published by Rose et al (128). Rose (128) has proposed a theory of sieving based upon the hypothesis that the kinetics are controlled by the blinding. The validity of this approach has been confirmed and the equations derived used to analyse continuous screening. However since these equations require the application of fairly complicated computational techniques to each specific problem, the use of this model is inappropriate for simulation purposes. Therefore Rose (128) himself has suggested the formulation of an empirical equation, in the form of a partition curve, to allow the calculation of the outlet streams:-

$$E = E_o + (E_\infty - E_o)(1 - e^{-(\delta t)^n})$$

where $E_o = e^{-KL/V}$

$$E_\infty = E_o^{(k/(k_1 + GkRP/V))}$$

$$n = E_o^{-0.127}$$

$$\delta = \frac{0.0132\ KGR}{K^{0.4}L^{0.4}V^{0.6}}\ e^{4.11(V/GR)^{0.5}(K^{0.15}k^{0.5}/k^{0.65})}$$

and K = Coefficient of diffusion of undersize material through sieve cloth (min^{-1})

L = Length of screen (m)

V = Rate of motion along screen (m/min)

k_1 = Coefficient of diffusion of blinding material out of sieve cloth (min^{-1})

k = Coefficient of diffusion of blinding material into sieve cloth (min^{-1})

R = Rate of feed per unit width of screen (Kg min^{-1}/m)

P = Mass of blinding material per unit mass of feed material

G = Area of sieve cloth blinded by unit mass of blinding material (r

The three parameters 'K', 'k' and 'k_1' are characteristic of a particular machine, whilst 'G' and 'P' are unique to each feed material, and so five parameters need to be measured by experiment for each screen application and variation in feed. Whilst this may be a useful basis for simulating operating plants and for future development of screening models, this parameter determination precludes the model's use for design purposes until methods for characterisation of industrial screens and process streams have been found.

A screening model used by Gottfried (129) in the simulation of coal preparation plants is based on a selectivity function, "$C_s(x)$", which describes the fraction of material of size "x" passing through a screen opening of size "y_o". It is stated that testwork shows thus function to be of the form:-

$$C_s(x) = \exp(A(1 - x/y_o))$$

where A is the separation strength.

This function describes a curve contrary to the shape of more commonly used partition curves. "Reasonable values" are used for the empirically determined separation strength which is dependent on screen type, screen opening, production rate, screen location and operating conditions. However few details of the magnitude of these values or how they may be determined are given and so their use, especially in metalliferrous plants, is somewhat risky.

Taking into consideration the need for screen models to incorporate the data collected over many years by equipment manufacturer's and for parameter definition to be possible without even preliminary testwork, Karra (130) has developed a screen performance model based on existing screen sizing methodology and data from a specially designed experimental campaign. Regression analysis was carried out on the data which Rexnord Inc. (131) use to predict the various empirical factors used to allow for feed stream variables. The screen partition curve can then be calculated using the empirical equation:-

$$\%C_i = 100\{1 - \exp(-0.693(d_i/d_{50})^{5.846})\}$$

$$\text{where } d_{so} = h_T \left(\frac{\text{Theoretical undersize (tph/sq.m)}}{\text{ABCDEFG}}\right)^{-0.148}$$

h_T = $(h+d)\cos\Theta - d$

and C_i = oversize partition coefficient for size fraction i

d_i = particle size, fraction i (mm)

A,B,C,D,E,F,G = Empirically determined factors

h = square mesh aperture (mm)

d = wire diameter (mm)

Θ = screen angle of inclination to horizontal.

This model allows a mass balance to be calculated for each size fraction in a stream and, whilst empirically based, does give a phenomenological insight into the significant variables in a screen application. It is also of a form that most minerals engineers can understand and easily relate to, and makes use of screen manufacturers' experience.

2.2.2 Design and Modelling of Hydrocyclones

Hydrocyclones are almost universally used for mill classification between 150 and 5 microns, although coarser separations are possible; whilst this is their main application, hydrocyclones are also used for de-sliming, de-gritting and de-watering (132). A typical hydrocyclone is shown in Fig. 2.6(a) and some of the important factors affecting performance have been reviewed recently (53, 132, 133); the internal flow regime appertaining under normal operating conditions is shown in Fig. 2.6(b). Particles larger than those desired in the overflow (vortex finder) should remain close to the wall and travel downwards in a spiral path and exit through the underflow (apex). Classification of solid particles of different weights occurs as a result of the conversion of the direction and velocity of the inlet flow into those of the outlet streams and is due to centrifugal drag and gravity forces acting on the particles; unlike screening, in hydrocyclone classification particle weight is the major separation factor rather than size.

Design of hydrocyclone installations is normally based on the study of manufacturers' charts which show capacity and separation size range in terms of hydrocyclone size; in recent years the trend has been to describe performance graphically for each narrow size fraction of feed solids. These graphs are in the form of corrected

partition curves (see Chapter 2.9) for which a number of empirical functional forms have been proposed. These techniques are generally found to be adequate, although care is required in their implementation, and no testwork is needed except perhaps for detailed design work or when a difficult ore is to be treated.

Therefore if hydrocyclone installation design is to be enhanced through the use of computer-aided techniques, a model which can accurately reflect the use of published partition curves and, if possible, give some insight into the effects of design variables is needed. Before 1964 there was a great deal of both fundamental and applied research into the operation and principles of the hydrocyclone. There were attempts to model the fluid flow in a hydrocyclone by a number of workers (134 - 136); one of the most significant research works on fluid flow remains that of Kelsall (134) who established the framework for much of the work which followed. In 1964, Bradley (137) reviewed the state of the art, listing eight different equations to calculate the cut size of a hydrocyclone and nine different equations to predict the pressure drop. Whilst some authors (53, 132, 138) have claimed that much of this work is not directly applicable to mineral industry circuits because the results were based on testwork using dilute slurries and small diameter hydrocyclones, Ford (16) has shown not only how similar the equations were to each other, but also how similar they are to recent models based on more suitable testwork.

One of the more recent models, that of Trawinski (139), is only a development on Bradley's (137) and Krijgsman's (140) work and suffers from the same potential limitations. Another model of restricted application is that of Mular and Jull (133), developed from testwork in Krebs cyclones (141). This is largely a simplification and formulation of the manufacturers' graphs and design procedures described earlier for 'typically' shaped hydrocyclones. Whilst of use in preliminary design when selecting hydrocyclones for a particular duty, this model is perhaps of a too basic and empirical nature for optimum use in process simulation for design purposes; also, whilst deriving equations for calculating the cut size and pressure drop, there is no formulation of the shape of the partition curve so that it is difficult to predict the effect of classification on a heterogeneous particle population.

Since Bradley's review (137), the most significant research on hydrocyclones has been carried out by Lynch, Rao et al (142 - 146) on industrial scale apparatus in Australia. Other models, notably Plitt's (138), and industrial simulations (13, 132, 147) have used this work as the basis for further developments and applications. A series of largely empirical equations, which are mechanistic in form and therefore useful in understanding the separation, have been developed to describe four fundamental parameters in terms of operating and design variables; these parameters are the classification size, partition curve, water flow ratio and pressure-throughput relationship. This model has proven to be useful in simulating operating plants (13, 132) although extrapolation from one set of conditions to another has been found to be erroneous (147); this is because certain empirical constants used in the model are specific to each hydrocyclone installation and must be determined by testwork. Whilst it is claimed (146) that these are constant over a wide range of flow rates and conditions, it has been shown (147) that the useful range is quite narrow. Therefore the use of the Lynch et al model is somewhat restrictive for design purposes.

Plitt (138) used the results from the work of Lynch et al and further testwork to develop another model, more theoretically based, designed to be universally applicable and not dependent on experimental work; the model is stated to be adequate for design purposes in that it defines the effect of hydrocyclone geometry and of operational variables on performance without the need for any additional experimental data, though it has been found that the model does not always accurately describe operating plants (147).

The model equations are:-

$$S = \frac{1.9(D_u/D_o)^{3.31} h^{0.54} (D_u^2 + D_o^2)^{0.36} \exp(0.0054\phi)}{H^{0.24} \quad D_c^{1.11}}$$

$$R_v = \frac{S}{S+1}$$

$$R_f = \frac{R_v - R_s\phi}{1 - \phi}$$

$$D_{50}(c) = \frac{50.5\ D_c^{0.46} D_i^{0.6} D_o^{1.21}\ \exp(0.063\emptyset)}{D_u^{0.71}\ h^{0.38}\ Q^{0.45}\ (\ell_s - \ell)^{0.5}}$$

$$Y' = 1 - \exp\left(0.693\left(\frac{d}{D_{(c)}}\right)^m\right)$$

$$m = 1.94\ \exp(-1.58\ Rv)\left(\frac{D_c^2 h}{Q}\right)^{0.15}$$

$$P = \frac{1.88\ Q^{1.78}\ \exp(0.0055\emptyset)}{D_o^{0.37}\ D_i^{0.94}\ h^{0.28}\ (D_u^2 + D_o^2)^{0.87}}$$

where $\emptyset$ = volumetric fraction of solids in feed

H = pressure drop across hydrocyclone (as head of feed slurry)

$= P\ \ell_p$

P = pressure drop across hydrocyclone (kilopascals)

R_f = water recovery to underflow

R_s = solids recovery to underflow

R_v = pulp recovery to underflow

$D_{50(c)}$ = corrected cut size (μm)

Q = volumetric flow rate of feed pulp (l/min)

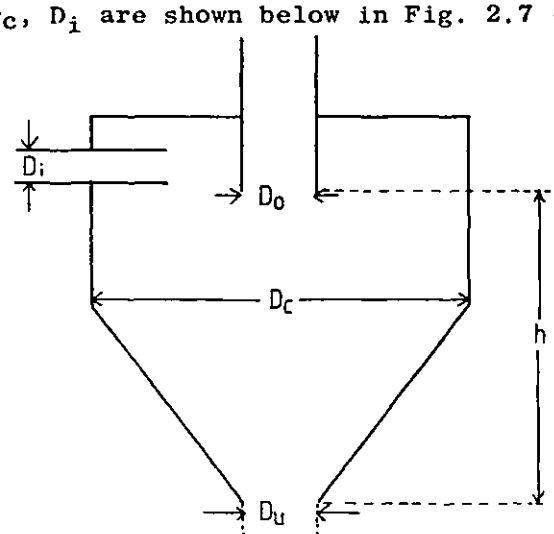
ℓ_s = solid density (g/cm^3)

ℓ_p = pulp density (g/cm^3)

ℓ = liquid density (g/cm^3)

d = particle diameter (μm)

D_u, D_o, h, D_c, D_i are shown below in Fig. 2.7 (cm)



and Y' describes the shape of the corrected partition curve.

Whilst the model predicts, with reasonable accuracy, the flowrates of solid particles around a hydrocyclone, prediction of the water flowrates around the hydrocyclone has so far defied mathematical treatment. Plitt has reviewed the literature on the split of water and pulp between underflow and overflow; most workers have used empirical relationships of the form:-

$$S = \left(\frac{D_u}{D_o}\right)^x$$

where S = underflow volumetric pulp flow rate / overflow volumetric pulp flow rate

x = constant.

The value of 'x' varies between workers from 3.0 to 4.4. Plitt's results indicating a value of 3.31 though the equation also includes the term "$(D_u^2 + D_o^2)^{0.36}$" and allows for the effects of free vortex height (h), feed pressure hydrocyclone diameter (D_c).

Whilst not necessarily more accurate than other models, the broad data base used by Plitt and the inclusion of the most significant operating variables allows use of the model in a design application for a wide range of operating conditions; also the form of the model allows the relative importance of major variables to be revealed. As and when more experimental data becomes available, appropriate constants in the model can easily be adjusted to improve the model as necessary.

One of the most restrictive problems in the use of this model in process simulation for design purposes lies in the numerical solution of the model equations and the necessity for so many variables to be defined. To improve computational efficiency Plitt's model has been adapted for simulating "typical" hydrocyclones (16) by making assumptions concerning the values of certain variables. The major assumption made is that the water recovery to the underflow is equal to the pulp recovery; whilst this is only approximately true for dilute slurries and even less so for higher solids concentrations, the assumption is no less appropriate or more restrictive than the various empirical relationships used in other models and has the advantage of removing the need to solve the equations iteratively. The other assumptions include defining a 'typical' hydrocyclone geometry and assuming that

the fluid medium is water. Making only these assumptions reduces the model equations to:-

$$S = \frac{0.323\ D_c^{-0.39}\ h^{0.54}\ \exp(0.0054\emptyset)}{H^{0.24}}$$

$$R_v = \frac{S}{S+1} = R_f$$

$$D_{50(c)} = 35.03\ D_c^{0.79}\ h^{-0.45}\ P^{-0.252}(\ell_s - 1)^{-0.5}\ \exp(0.063\emptyset)/\exp(0.0055$$

$$Y' = 1 - \exp\left(0.693 \left(\frac{d}{D_{50(c)}}\right)^m\right)$$

$$m = 2.94\ D_c^{0.044}\ h^{0.127}\ P^{-0.084}\ \exp(-1.58\ R_v)$$

A "typical" hydrocyclone is defined as having dimensions (139, 141):-

$$D_o = 0.1666\ D_c$$
$$D_i = 0.20\ D_c$$
$$D_u = 0.116\ D_c$$

2.2.3 Summary and Conclusions

Classification of mineral particles according to size is largely by means of vibrating screens for coarse size separations and hydrocyclones over the finer size ranges. Scale-up of neither of these machine types is accurate from the results of small scale testwork and so preliminary design is usually dependent on partition curves derived from equipment manufacturers' data bases. Therefore any useful simulation model for design purposes should be independent of testwork for parameter determination and be capable of predicting the partition curve for a size separation installation under varying feed conditions.

Whilst the approaches used by Ferrara et al (125) and Rose (128) offer the basis for more theoretically based screening design procedures, they still require further development, preferably utilizing manufacturers' data. The most appropriate model, implicitly using this data, has been developed by Karra (130) based on Rexnord design methods (131). The model uses an empirical relationship to describe the screen partition curve; the actual shape of the curve is based on analysis of the feed material and is adjusted using equations derived from Rexnord design factors.

The major developments in hydrocyclone modelling appropriate to the design of mineral processing plants have been based on testwork carried out by Lynch et al (146). For repetitive and rigorous use in a process simulator, a modified version of the Plitt model (138) is most suitable; this has used further testwork to augment Lynch's data and can be used for design with no extra experimentation. The large number of parameters to be determined and the iterative nature of the model solution procedure suggest that a simpler and more robust form of the model is more appropriate than the original; however no testwork has been carried out to test the validity of the simplifying assumptions.

2.3 Solid-Liquid Separation

Solid-liquid separation can be one of the most capital intensive areas in a minerals extraction system and is also where many of the major problems are encountered in wet treatment processes. The aim of the separation is one of:-

1) Solids recovery
2) Liquid recovery
3) Recovery of both the solid and the liquid
4) Recovery of neither solid nor liquid

Due to the high costs involved, solid-liquid separation has been the target of a great deal of research, both directly, to improve the efficiency of the techniques used and the understanding of the basic mechanisms, and indirectly, to develop new extraction processes to eliminate the need for such operations; however these latter developments are still in their infancy and so the economics of solid-liquid separations are often of vital importance in the evaluation of a proposed process.

There are numerous methods available for carrying out these separations, including many forms of filtration, centrifugation, sedimentation, flotation, cycloning and screening, and aids, such as flocculants and filter aids, for improving their efficiency; indeed the number of satisfactory and alternative choices is so large as to preclude the choice of a best solution (148). In the selection and definition of a separation system for a particular situation, theory is used only in the crudest form, as the complexity of solid-liquid separation phenomena frequently renders mathematical analysis inadequate; therefore bench scale testing and experience play predominant roles in equipment design. Over recent years, various attempts (148 -150) have been made to tabulate the key factors relevant to solid/liquid separation equipment, so as to provide a kind of "rapid selection guide", although the sheer number and diversity of factors make this a difficult objective; these methods tend to specify equipment in terms of process scale, mode of operation and objective. Solid-liquid separation processes are usually grouped into two broad categories: sedimentation and flotation, and filtration. Based on this categorization a preliminary selection programme (149) might comprise four stages:-

1) Definition of separation duty.
2) Quantifying slurry settling characteristics by basic testing.
3) Quantifying slurry filtration characteristics by further testing.
4) Comparison of problem specifications 1 - 3 with the defined characteristics of equipment types.

In the minerals industries, the selection procedure is simplified somewhat if only those processes traditionally used are considered; although rake and spiral classifiers, centrifuges and many forms of filtration are excluded, modern, high tonnage metallurgical operations generally rely on hydrocyclones, thickeners and/or vacuum filtration.

Before studying the design of thickeners and filters in detail, it is worth noting that in general design methods for solid-liquid separation equipment are solidly founded on substantial industrial experience, and yet none merits worth deeming as a standard procedure, this total absence being somewhat remarkable (151). In some sectors a relative uniformity of practice can be seen, but elsewhere there are wide differences in the techniques and procedures used by the suppliers of very similar equipment types. Also, in general, knowledge and experience of this equipment are concentrated in the hands of manufacturers rather than in those of the equipment users; it is therefore normal to have to consult these vendors when a new installation is under consideration.

In common with other unit processes, solid-liquid separation is never perfect; some solids invariably leave with the liquid stream and some of the liquid will always leave with the solids laden stream. The most common way of describing separation efficiency is to consider the solids and liquids separately so that two independent factors can then be defined; these are the mass fraction of solids recovered and the moisture content of the solids stream. For the solids stream it is normal to define an overall recovery factor, thus treating the particle population as homogeneous. As will be seen there is an almost total lack of detailed process models, and mass balances around a flowsheet are usually calculated based on the definition of these two factors.

2.3.1 Thickeners

The most commonly used thickener in minerals operations is of the circular-basin type shown in Fig. 2.8, although plate or Lamella thickeners are gaining rapidly in popularity; descriptions of thickener machines and operations can be found in the literature (e.g. 152-159). Although labour and power costs for these devices are not negligible, their over-riding economic consideration is their capital cost; whilst they are high throughput operations and therefore residence time and volume capacity are important in the final design, thickeners can generally be costed in terms of their horizontal area. Due to the consequent importance of thickener area, a great deal of research has been carried out to develop reliable methods for its prediction.

This very marked attention to sedimentation theory and design has not yet been reflected by the current state of the art, and there is still not a clear design method by which design engineers may arrive confidently at an optimum equipment size based on batch settling tests; numerous writers (152, 157, 158, 160-164) have commented that the available models are not uniformly valid and in some aspects quite dubious. Whilst models for determining thickener area from batch settling tests have been proposed, many being the basis for industrial design methods despite the doubts about their theoretical validity, thickener sizing is still very much of an art and adequate designs are only obtained through the use of "efficiency" or "safety" factors. Even if experienced judgement and discrimination are applied to the results of laboratory tests, it is still necessary to carry out pilot plant work for reliable results and designs.

Although the theoretical nature of some of the models is responsible for these drawbacks, the basic failure in predicting thickener design requirements from laboratory testing is that the standard batch settling test does not replicate precisely the conditions found in a commercial thickener (160). If settling rate were a unique function of concentration, then static bench tests would be satisfactory, but other factors such as pulp height, weight of solids/unit area and the pressure of raking, affect the velocity for a given concentration; also consideration of the construction and mode of

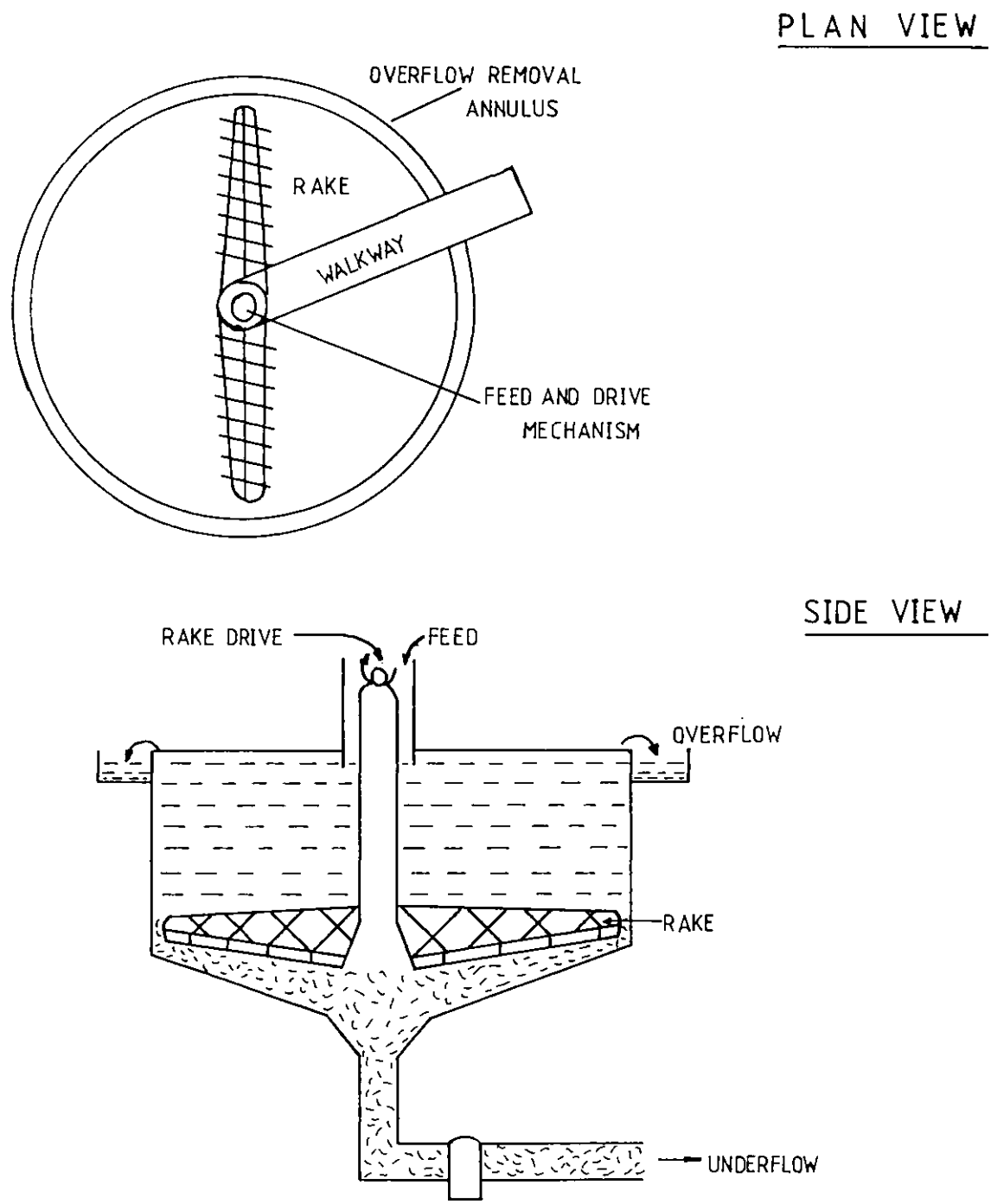


Figure 2.8 Schematic representation of a thickener

operation of a commercial thickener highlight the differences likely to be experienced that are not simulated in a simple batch tests.

Since the publication of Coe and Clevenger's (165) classical paper it has generally been considered that the concentration of a slurry occurs in one of two basic regimes; free settling or compression. The boundary between the two is the transition zone, which is at the "critical concentration". Above this critical concentration the slurry is in a compressed state and below it in free settling conditions; the zones that Coe and Clevenger observed are shown in Fig. 2.9. They assumed that in the free settling zone, the settling

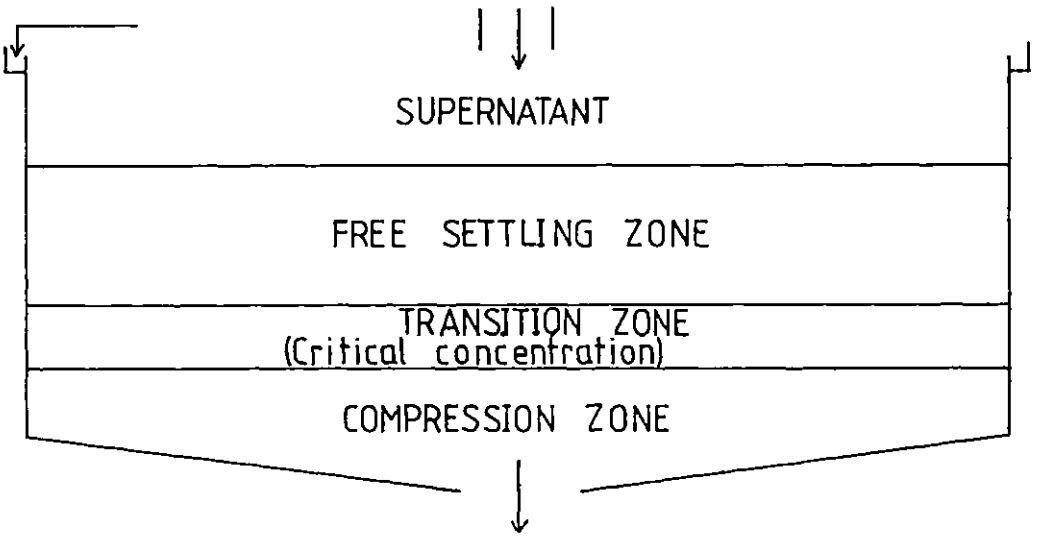


Fig. 2.9: Coe and Clevenger thickener zones

velocity of the particles is a function only of the solids concentration and also that this zone was controlling with respect to area requirements; these assumptions formed the basis of their method for the calculation of thickener area and also the methods proposed by Kynch (166), Talmage and Fitch (167), and other similar models. The various methods and models have been comprehensively reviewed and discussed by Fitch (157) and Pearse (160), amongst others; there have been a number of attempts (168-170) to compare these design methods through direct testing, but the results of these comparisons have been somewhat contradictory, highlighting perhaps the limitations of these methods. It should however be noted that whilst not theoretically sound and needing empirical safety factors, these methods have been found through

a great deal of practical experience to provide a useful tool for obtaining acceptable preliminary estimates of thickener dimensions from batch settling tests.

Other than the problems already discussed, such as the nature of batch testing and the dependence of settling rate on other factors than solids concentration, a major drawback of the Coe and Clevenger type of approach is their lack of application to compressible pulps, such as exist in flocculated slurries. For compressible pulps the theories of Kos (161, 171) and Adorjan (172, 173) offer the potential for progress but some testing and apparatus simplification need to be done before their reliability and limitations are known. With the increased use of flocculants this will become an even more important area for development.

Considering this lack of suitable development in the field of thickener design, it is hardly surprising that it would appear that there has been only one attempt (174, 175) to develop even a phenomenological model of the continuous operation of a thickener based on the results of small scale testwork and that this model should be based on one of the design methods developed from Coe and Clevenger's work. In this model Kynch's theory (166) is superimposed on a plug flow fluid pattern to allow the effects of the depth of sludge blanket and solids residence time on thickener capacity to be estimated; then for a given maximum residence time and total solids underflow, an optimum settler design is arrived at by balancing area, height and underflow concentration so as to minimize cost. Considering the model's complexity and dubious inherent assumptions, it would appear to offer few advantages over existing mass balance techniques which are to assume values for the mass fraction of solids recovered, and the moisture content, in the underflow stream; a suitable thickener area and depth are then chosen based on the throughput such that these values can be maintained in conjunction with a clear overflow stream.

2.3.2 Vacuum Filters

For dewatering operations rotary drum and disc filters are most commonly used, whilst for processes where cake washing to remove surface liquor is necessary, the previous predominance of drum filters has recently been challenged by the introduction of large and efficient horizontal belt filters; descriptions of these filters and their

operation can be found in the literature (e.g. 176-180). Although the pressure drop, and therefore vacuum pump capacity, is important to both the capital and the operating cost of a filter operation, and the labour and power costs are also not negligible, the design and evaluation of a filter operation is very similar to that of a thickener operation, in that filter area is perhaps the major and over-riding economic factor. Another similarity is that process mass balances are generally calculated using a priori assumptions of separation efficiencies, treating solids streams as homogeneous, and that theoretically based models are only used to determine pressure drop, filter area, and, only in a few cases, washing characteristics rather than acting as simulation models.

Design techniques for filters are also similar to those for thickeners, in that initial, approximate estimates are made from small scale tests, which do however more closely simulate industrial operations than their counterparts, but that accurate design work can only be carried out following pilot scale work. Due to the cyclic nature of continuous filter operations with cake formation, cake dewatering, cake discharge and, possibly, cake washing sections neccesary, it is essential to simulate each of these in testing as one of these will be controlling in terms of area requirements and they must all be considered in order to incorporate them into the filter cycle. This testing is normally carried out using small test leaf filters of area 0.01 m^2 or 0.1 ft^2 linked to laboratory vacuum systems, preferably capable of measuring the applied vacuum; detailed descriptions of experimental procedures have been provided by Osborne (181), Dahlstrom and Silverblatt (176) and Cain (177).

There are many phenomenological models of varying complexity and empiricism to describe the three rate functions in a continuous filter cycle (176, 177, 180-188), although few of these have been rigorously tested against industrial operations; most are of similar form to those proposed by Dahlstrom (189):-

Cake Dewatering Rate: $\%M = f(Fa, d, \%Mr)$

$$Fa = f\left(\frac{\Delta P}{W} \cdot \left[\frac{CFM}{Ft^2}\right]_d \Theta_d\right)$$

Cake Washing Rate: $\Theta_w = K' . \Theta_{f.n}$

$$R = 100\left[1 - \frac{E}{100}\right]^{n}$$

Cake Formation Rate: $$Z = \left[\frac{K.w.\Delta P}{\mu.\alpha.\Theta_f}\right]^{\frac{1}{2}}$$

where

%M	=	weight % of moisture of the discharged cake
F_a	=	approach factor depicting approach to %Mr
%Mr	=	weight % of residual moisture
d	=	particle size, particle size distribution and shape factor
ΔP	=	pressure drop across the cake
W	=	cake weight of dry solids per unit area per cycle
$\left[\frac{CFM}{Ft^2}\right]_d$	=	gas pulled through the cake measured at down stream pressure as cubic feed per minute per square foot during the dewatering portion of the cycle
Θ_d	=	dewatering time during the cycle
Θ_w	=	wash time during the cycle
Θ_f	=	cake formation time per cycle
n	=	volume of wash fluid/volume of cake liquor
R	=	% of soluble solids remaining in the cake after washing
E	=	wash efficiency
Z	=	weight of dry solids per unit area per unit time of cake formation
w	=	weight of dry cake solids per unit volume of filtrate
μ	=	viscosity of liquid
α	=	cake resistance
K,K'	=	proportionality constants

Some individual parameters such as K, K', α and E must be determined for each slurry system, and in the case of α for each filter media, and curves of %M and F_a plotted also for each slurry system. Even when the necessary tests have been carried out the results are not totally reliable due to the inaccuracies and difficulties involved with the testing system, and predictions are obtained for only the total solid and liquid mass flows only if filter area and pressure drop are defined. For a preliminary evaluation and mass balance it is perhaps best if estimates of the necessary area and pressure drop

are made from a simplified test campaign, and predictions of separation and washing efficiencies are made from experience; reliable designs and parameter determination can only be made based on pilot plant work, and so basic mass separation models should be used for simulation purposes and cost estimates should be based on a minimum of testing.

It is recognized (189, 190) that whilst both filtration theory and equipment appear to be basically simple, paradoxically their application to practical filtration is the source of immense confusion and difficulty. In order to standardize filter design procedures and to weld together the existing mass of fundamental knowledge, empirical experience and equipment based data into one useable, practical whole, proposals have been made for unifying cake filtration tests based on the concept of Standard Cake Formation Time (191); this would only involve measuring cake thickness at various cake formation times at a standard pressure drop across the cake. Whilst it would appear that this proposal may well be receiving the support it needs (190), further development is necessary to obtain the relationships between the Standard Cake Formation Time and slurry and operating parameters for various types of equipment.

2.3.3 Slurry Washing

An increasingly important application of solid-liquid separation is in hydrometallurgical processes where all of the soluble material needs to be separated from the insoluble material following leaching. As has been stated, few separations are perfect and so it is common practice to wash the solids bearing stream to remove liquor still being carried. Both filters and thickeners are used in various combinations for this type of operation according to local circumstances; depending on the quantity and value of the leach liquor, there may well be more than one washing stage, these stages often being arranged in a counter-current fashion.

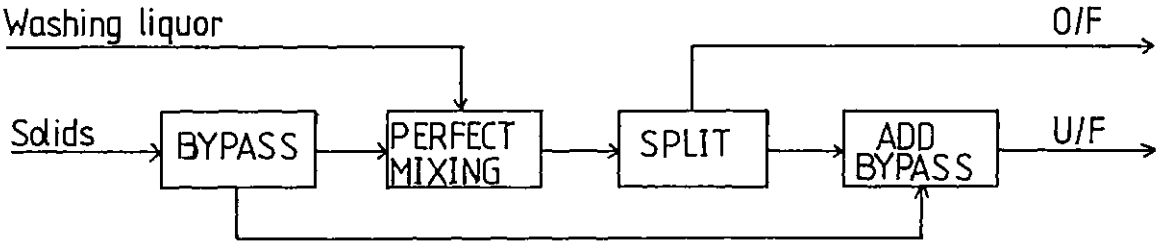
From a survey of the literature it would appear that no work has been published on the analysis of this application of thickeners, although there has been more extensive study of filter cake washing characteristics; this latter work has been reviewed by Wakeman (192) and the basis for a design method suggested. However this design

method is reliant on similar testwork to that employed for the more detailed filtration models and thus suffers from similar limitations.

Perhaps a more appropriate approach for filtration applications, as suggested by Tomiak (193-195) and Hermia (196), is for the design engineer, based on experience and preliminary testwork, to predict liquor hold-up analagous to washing efficiency, for a given filter unit area and to then treat the mass balance as a simple separation situation. This is a similar technique to that used in alumina extraction from bauxite (197, 198) when thickeners are generally employed; washing operations are represented by four steps:-

1) Bypass of a portion of the solids inlet stream to the underflow without mixing with washing liquor.
2) Perfect mixing of the washing liquor and non-by pass portion of the solids inlet stream.
3) Splitting of the liquor for the overflow stream.
4) Adding of the remaining slurry with the by pass stream.

A schematic diagram of these four steps is shown in Fig. 2.10:-



The key parameter of this model is the mixing efficiency of the washing liquor and the solids inlet stream, which controls the amount of by pass and hence, defines the concentration difference between underflow and overflow liquid streams; perfect mixing would result in equal stream concentrations. If thickeners and filters are to be simulated through separation efficiencies and with an a priori knowledge of unit area, then it would appear logical to simulate all liquid-solid separation operations, including washing, using a similar and single model.

2.3.4 Summary

There are a number of applications for solid-liquid separation operations and a wide variety of methods by which these separations

might be achieved; however in the metals extraction industry most of these are carried out by sedimentation in thickeners or by filtration on a vacuum filter. Design techniques for these two types of machine are fairly similar in that initial requirements are estimated from the results of bench scale experimentation and a fairly large input of accumulated empirical data, but for detailed design work pilot-scale testing is required.

Whilst a great deal of experience has been obtained in the operation of these two processes, there still do not exist widely accepted approaches to the design of either thickeners or filters. There is still no theoretically sound or empirically proven way of predicting solid-liquid separation performance completely and reliably from measurements made on small samples of suspension (163).

In recent years there has been appreciable progress made in the understanding of fundamental models. However there is still the need for this work to be correlated with tests on industrial thickeners and filters so that the appropriate models can be developed into workable design equations for various types of machines.

Until such work has been completed it is likely that evaluation of these processes during the early stages of a project, will rely on overdesign of separation area to ensure adequate performance, and on prediction of separation and, in the case of washing, mixing efficiencies so as to obtain a process mass balance.

2.4 Froth Flotation

Whilst perhaps not the critical cost factor in extraction processes, froth flotation is the key metallurgical factor for most base metal and many precious metal and non-metallic operations; in the United States alone, in 1975 about 30 different mineral bearing materials, totalling nearly 423 million tons, were treated using flotation techniques (199).

2.4.1 Flotation Plant Design

Despite its age and pre-eminence as a concentration process, flotation is still largely an art whose success is 'a triumph of enlightened "know-how" over inadequate "know why". Progress has been made by empiricism tempered by engineering judgement' (200).

Fig. 2.11 shows the many complex microprocesses and numerous variables involved in the flotation process, each of which makes its own demands, so that the final process design is of necessity a compromise tailored in practice by engineering and economic considerations. Mika and Fuerstenau (201), drawing upon some 130 contributions to the literature, have described a model of the flotation process that is based upon the various transfer paths available within a cell and yet still failing to provide a reasonable basis from which flotation performance predictions can be made. Although the means by which solids are transferred from the pulp in a flotation cell to the froth concentrate are easy enough to grasp conceptually, detailed descriptions, qualitative or quantitative, of these transfer processes are not yet available. Since an intimate knowledge of the relationships and interactions among the variables is a pre-requisite for the success and profitability of a process, it is therefore essential that flotation testing is planned by experienced engineers, with perhaps less consideration to theory than most other metallurgical processes require; similarly, although much of the work can be carried out by technicians, it is often important for engineers to be present to observe characteristics such as froth colour and quality, rather than relying solely on inanimate product weights and assays (58).

This lack of understanding of the basic processes is at least partly responsible for the great variety of cell types that have been

developed over the years (202); even ageing standard texts such as Taggart (78) and Pryor (203) describe well over 40 different machine designs. Some of these differences are not just subtle variations in the basic hardware but gross variations in the design principles; however the designs used in metals extraction processes are predominantly of the sub-aeration type, though a few pneumatic designs have recently been introduced for specialized purposes (200, 202). The several sub-aeration machines commonly used appear to be of similar metallurgical performance (200), indicating that the design differences are not critical in the flotation process or that they can be negated by adjustment of the operating parameters.

The range of tests available to flotation engineers during design is almost endless, as are the books and articles written about them; MacDonald and Brison (204) include a good summary of reference articles in addition to describing their own methods, and the recent volume, "Mineral Processing Plant Design", has a number of papers relating to modern industrial practice (58, 205, 206). There have been shown to be correlations in the performance of laboratory scale equipment to larger machines (208), and so there is room for the use of all the available methods, especially in the initial exploratory work.

In describing the testwork involved in the various stages of plant design, Arbiter et al (209) outline 5 levels of accuracy of study; although, as is common, confusing nomenclature is used, it would appear that stages II and III of Arbiter's scheme are commensurate with the adopted level C study (± 25% accuracy) of the proposed methodology (see Chapter 1.3).

In this study, enough work should be carried out to be able to assess the main alternative circuits and their possible effects on project viability, as well as establishing basic relationships between capacity, grind and flotation results. Although Arbiter (209) and King (210) indicate that pilot plant testing is a necessary step in plant design, it is common for this scale of work to be falling out of favour with preference given to more accurate laboratory work in batch and continuous cells.

One of the first tasks in such a flotation testwork programme will be to establish the optimum types and levels of reagent. This

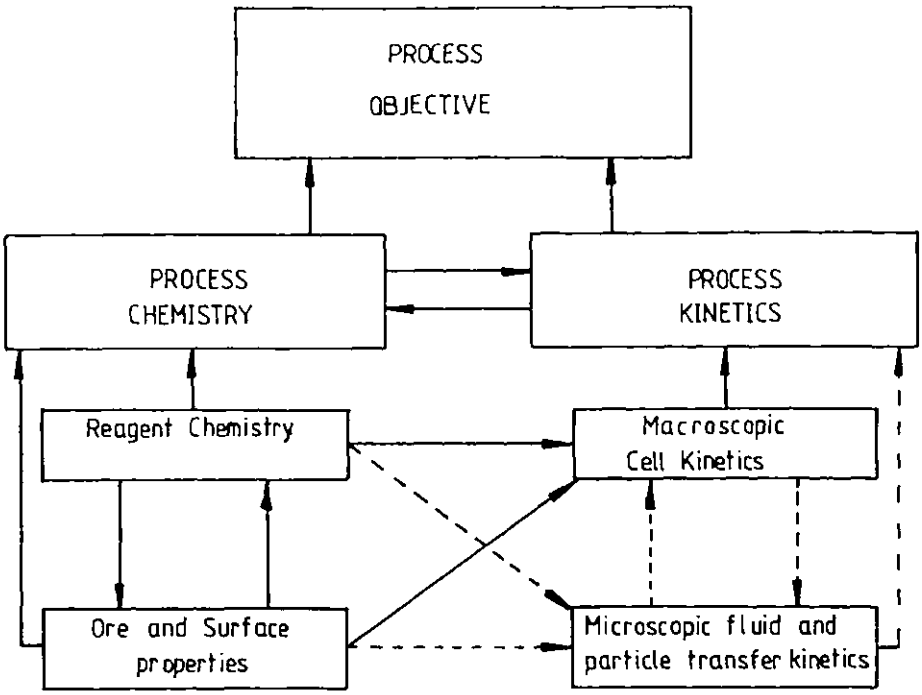


Figure 2.11 Summary of the various facets of a flotation problem (202)

(broken lines represent particularly poorly understood relationships)

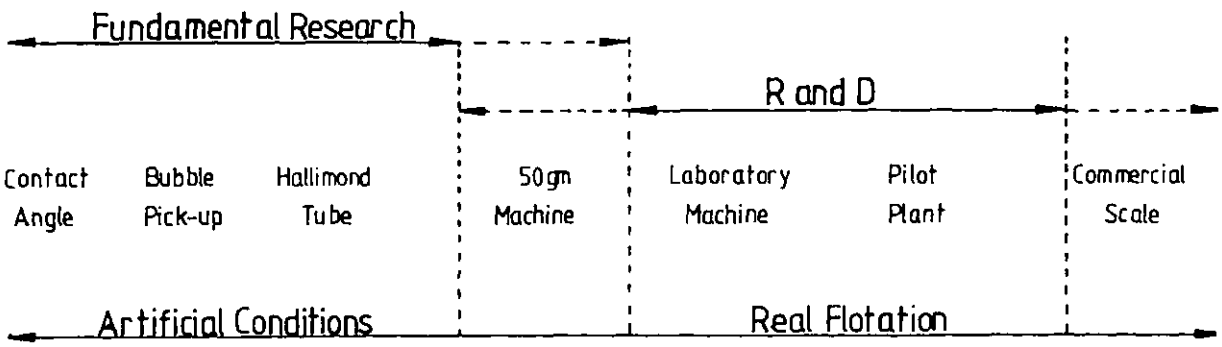


Figure 2.12 Harris spectrum of testing methods (207)

will depend, almost entirely, on the ore mineralogy, but despite the use of statistically designed tests and basic engineering models (206) in assessing the effects of changes, there is no textbook route to success and no substitute for an experienced engineer. There is not yet any quantitative measure of the effect of reagent mixes and so the investigator must rely on the results of previous similar testing along with a measure of intuitive guesswork.

Although the result of this work may indicate the probable flotation route, it will still be necessary to determine the quantitative effects of the plant configuration on recovery and concentrate grade. This is often initially carried out in batch flotation cells, with confirmatory cycle testing to follow when recycling is involved. Cycle tests tend to be long, tedious and difficult to carry out; there is no guarantee that a steady state will be reached in a reasonable number of cycles or that when steady state is reached the results will be satisfactory (211, 212).

Once the circuit configuration has been characterized it is necessary to determine the size of equipment required to produce a satisfactory performance and handle a stipulated tonnage throughput. This is often calculated from results of laboratory batch tests, which require scale-up by factors due to equal retention time, or, as is preferable, from continuous testwork which allows a more realistic cell residence time to be used directly in scaling up. With the knowledge of the time needed to obtain the desired recovery, flowrates (including air) and pulp density, it is possible to calculate the necessary cell and bank volume. This may be through the use of empirically derived kinetic curves describing overall recovery (213) or through simple models which may be used to describe the flotation behaviour of different fractions of a plant feed. Whilst the latter technique may require more testwork to determine individual rate coefficients, the extra amount is not excessive and the possible increase in accuracy and information that this approach could yield should outweigh the extra costs.

2.4.2 Modelling of Steady-State Flotation

There have been many reports of the development of mathematical models intended to describe the flotation process, these having been comprehensively reviewed by Cooper (214), Woodburn et al (215, 216),

Harris (217) and Herbst and Mular (13). The published models used vary, according to amount of included detail and application, from quite complex largely mechanistic models, such as King (218), to simple regression models of plant performance, such as Dorenfeld and Beebe (219). Although many investigators have developed models in order to simulate a flotation plant, the number of such simulations published is, however, fairly limited.

Totally mechanistic models, as developed by Mika and Fuerstenau (201), are of no industrial application due to the many variables whose effects are unknown; similarly, totally empirical models are of limited value due to the inability to extrapolate beyond a model's range of application. Therefore the generally applicable models are of a semi-empirical, semi-mechanistic form, the two most commonly used approaches being the probabilistic and chemical kinetic analogy techniques. These are very similar in form, but a distinction is worthwhile to preserve the insight that the probabilistic approach gives to the process of flotation (13).

Gaudin (220) first proposed a hypothesis relating bubble-particle interaction which was then used by Schumann (221) to assess the probability of a particle being successfully floated. Flotation of a particle depends on the probabilities of (220):-

(i) a bubble-particle collision (Pc)

(ii) the collided particle adhering (Pa)

(iii) the particle then being retained and lifted through the pulp (Ps)

(iv) the particle being retained in the froth (Pf)

Therefore:-

$$R(\text{overall}) = Pc.\ Pa.\ Ps.\ Pf$$

Once a relationship for the joint probability is established, it is relatively simple, using population balance models, to calculate mass rates of flotation. Kelsall (223) combined the different mechanisms into a composite probability of flotation and applied this to discrete fractions of narrow size and mineralogical range to compute the probability of feed particles leaving in the concentrate. He also showed that this treatment was equivalent to a simple 1st-order rate model. Kelsall formulated his model as:-

$$W = Wo\ (1 - Poverall)$$

where Wo = weight of a component in feed stream

W = weight of a component in tailing stream

The description of flotation by analogy with chemical kinetics can be traced to Garcia-Zuniga's work in 1935 (224) and, although strictly an empirical method, has since been used extensively by other workers in this field. The apparent order of the rate equation is taken to be unity in a majority of the models though this has been the source for much controversy. Notably this concerns Arbiter's contention that the published data was consistent with a second order model (225); this was later repudiated by Morris (226) and Brown and Smith (227). However these arguments have only served to highlight the lack of theoretical and physical basis for the analogy.

In general, the flotation rate coefficient can be considered as a product of the probabilities described by Tomlinson and Fleming (222) and it can be shown that most published models are of the same form though differing in approach and detail. The formulation generally used is:-

$C_i = K_i \cdot F_i$ where C_i represents concentrate mass flow in the ith fraction

F_i represents feed mass in ith fraction

K_i represents the rate coefficient.

The main difference in these models is the method used to describe the distribution of the rate coefficient, K_i, when describing the flotation of a heterogenous particle population; the analogy of first order kinetics is reasonably accurate only for populations of particles of similar size and mineralogical composition. As it is reasonable to assume that every source of variation in the rate coefficient cannot be accurately described, it would appear natural to free the variation of the coefficient from any direct constraints and to postulate instead a uni-dimensional rate constant. This is the approach originally used, with individual rate constants applied to discrete particle classes of narrow size.

If one was to consider the extreme case of assigning each particle a rate constant, then with a very large particle population it is reasonable to postulate a distributed rate coefficient. This concept was first proposed by Imaizumi and Inoue (228) and was

extended and used by many other workers; two of the common distributions are the gamma (229, 230) and rectangular (231). Zaidenberg et al (232) proposed a 3 dimensional distribution as a function of population variables known to influence flotation rate. King et al (233) used this concept in terms of particle size, particle mineral composition and surface activity. A similar approach was used by Kelsall (223) who subdivided each size class into 2 sub-classes characterized as either fast or slow floating. However although the use of a particular distributed rate constant has been verified for some ores, there is no agreement on a universally applicable form for the distribution and so the use of discrete functions is less restrictive.

Some workers have found a need to allow for variation in rate constant within a size and grade interval. Loveday and Marchant (234) described this using a gamma distribution within each fraction, as did King (233) in his original model; the latter, although describing a population using continuous size and grade distributions, calculated using discrete classes and used the gamma function to describe variation within these classes. Later workers have tended to use only discrete classes, both to describe flotation distribution within a given class, but also to describe the size and mineral composition distribution within a population. King (235) later used this concept, describing variation of flotation rate within a particular class using a fixed discrete distribution. Lynch et al (236) and Sutherland (237) used only two classes of fast and slow floating species whilst Jowett (238) suggested a three class distribution with fast, slow and not floatable components. It would however appear that workers who have needed to account for flotation rate variation within a given class have used few of these classes to describe a very heterogeneous particle population, whilst those who have used more narrowly defined particle classes (239) have found it unnecessary to allow for this extra variation.

As was stated earlier, the flotation rate coefficient can be considered as a lumped parameter combining the probabilities of the possible mechanisms involved in the flotation of a particle. However, as Bull (240) found, the use of the rate coefficient alone can produce inaccuracies in scale-up. It is generally agreed that the description of the effect of the froth phase is the main deficiency in mathematical

descriptions of the flotation process.

Most attempts to take into consideration froth effects have involved modelling the actions of the froth and the pulp separately; Harris (217) has reviewed the proposed models and Cutting et al (241) have tested some of their inherent assumptions. From this work it would appear that, despite the proliferation of model structures and parameters, froth structure and froth residence time effects cannot be adequately characterized such that accurate predictions of plant performance can be made from laboratory test data (241). Even the flow characteristics of the froth could not be accurately described; Moys (242) developed the plug flow concept introduced by Cooper (243), whilst the commonly preferred model of Harris and Rimmer (244) represents the froth as perfectly mixed. Cutting et al's (241) work showed both models to be inappropriate.

The NIM[1] simulator (210, 235, 245-247) uses a froth transmission factor to estimate the proportion of froth produced that passes over unit length of the cell lip into the launder; Sutherland (237) uses a similar concept utilizing a maximum concentrate flow rate beyond which the flow rate is restricted by a totally empirical function. Although perhaps providing adequate qualitative descriptions, quantitatively this approach relies a great deal on the judgement of the user especially when predicting plant performance from laboratory data.

One of the major assumptions often made in flotation modelling is that concerning the effect of pulp density on rate. As shown by a number of investigators (222, 228, 248), flotation rate is largely independent on pulp density for dilute solutions but becomes dependent on aeration rate above 60% solids; this gives rise to the concept of 'free' and 'hindered' or 'inhibited' floating conditions. Recent plant testwork by Frew (249) indicated a possible correlation between pulp density and flotation rates in a cleaner circuit over the 30-50% solids range; however a dearth of data precluded any claim of generality. For reasons of simplicity therefore, modellers have tended to assume that the former condition prevails, a reasonable assumption in many cases. King's model (245) allows for aeration rate

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[1] NIM = National Institute for Metallurgy (now Mintek), South Africa.

2.4.3 Simulation of Flotation Plants

Despite the work that has been carried out on flotation modelling, usually with the intention of simulating the process, there have been few reports of successful plant simulations. Sutherland et al (237, 253, 254) have used simple flotation models in a number of studies of the arrangement of flotation cells, but these have never been correlated against actual plant data to check their validity.

The most readily available and publicized general purpose flotation simulator is that developed at NIM by King et al (210, 235, 245-247). Extensive work has been carried out in assessing its use in scale-up from laboratory data to larger scale continuous plants (255); although failing to meet 95% confidence limits, the results were accurate enough to be used in meaningful studies though predictions of concentrate grade should be treated with some caution.

Stewart et al (239) also successfully used simple flotation models in the framework of a plant simulator to study alternative circuit arrangements; model parameters were, however, obtained from plant data and predictions from the model of the effect of changes to the plant were not compared to the actual changes in plant behaviour. Loveday and Marchant (234) used similar techniques in their study of a copper concentrator but again there was no verification of the model predictions.

A recent study by Frew and Trahar (256) is one of the few published comparisons of actual to predicted plant performance. Their results indicate that whilst reasonable predictions can be made from laboratory data, differences in flotation rates occur between sections in a plant even if the physical environment remains the same.

Three main approaches have been made to modelling the behaviour of the different sections:-

1) assume rate coefficients are the same in each section of a plant (228)
2) physical factors such as aeration rate, agitation and pulp density have an effect on flotation rates and change throughout a plant; the model used allows for such changes (234, 235)
3) different rate coefficients are measured for each section.

The work involved increases from approach (1) to (3) as does the accuracy; however the use of scaling factors applied to the rate

coefficients allows approach (3) to be used but relies on the assumption that changes act uniformly on all sizes and species.

2.4.4 Summary and Conclusions

In general the overall conclusion is reached, although total mechanisation of process performance prediction cannot be achieved readily, either now or in the forseeable future, that predictive modelling techniques can assist in flotation process design. However, to have industrial application, computer aided design must combine flotation experience, laboratory test data experience and predictive modelling techniques.

Generally, design of flotation plants uses laboratory scale testwork to derive empirical relationships of recovery and grade, which in turn are used to calculate plant size. However the effect of recycle streams and therefore cleaner performance is very difficult to determine without the use of models and simulation and so large margins of error are allowed for.

Semi-mechanistic models are becoming more and more commonly used though there are few published plant simulations. Most engineers accept that a plant simulator would be useful and that the testwork involved in determining model parameters fits in with normal design procedures.

State-of-the-art models are of the form

$$C_i = K_i \cdot F_i$$

The equation is best applied to narrowly defined, discrete fractions of the particle population as the continuous distributions proposed are not of general application. The fractions are often further subdivided, nominally to allow for changes in surface properties and mineral composition; however it is likely that if the fractions are defined narrowly enough then further subdivision is superfluous. It is also likely that this technique may be used to hide other deficiencies of model structure; if the model is kept simple it will tend to be more flexible and robust.

It is however impracticable to allow for changes in reagent regime and other physical variables such as impellor speed, pulp density and aeration rate. The best approach is to attempt to keep laboratory conditions similar to plant conditions and determine the possible reagent schemes through an organised laboratory test programme;

parameters for the use of the model can then be determined. Rate coefficients should, where possible, be measured for each plant section or in the absence of data, engineering judgement should be used in scaling the coefficients. It should be remembered that the rate coefficient is a lumped parameter and that, for design purposes, it is best not to clutter the model with too many other parameters.

The effect of froth action is poorly understood and no accurate models exist. For plant design purposes especially, control of total concentrate flow should be based largely on experience and to treat flotation as a two-phase problem rather than three phase. A similar situation exists for determining water flows; again judgement should be exercised based on experience, probably through the control of an exit stream pulp density.

In the modelling of flotation cells, it is both common and reasonable to assume perfect mixing of the pulp. However this is in theory incorrect and so, especially with the introduction of larger flotation machines, it is important that more characterisation of industrial cells is carried out.

2.5 Leaching Copper Ores

Although occasionally, but increasingly, applied to the treatment of flotation concentrates, hydrometallurgical extraction techniques are mainly used in the treatment of oxidized copper ores and to low grade oxide and sulphide mine wastes. Approximately 900,000 tonnes of copper were produced in 1974 using leaching, representing some 15% of the world's primary copper production (257). Whilst almost all of the techniques and approaches used in copper leaching are very similar to those practiced in the leaching of other metals, the analysis of each solid-liquid system is unique in its details, the mechanisms involved and the system's chemistry, requiring, for the present, concentration on the leaching of copper; it should, however, be noted that much of the discussion and the broad conclusions made are likely to be appropriate for other leaching systems.

When studying the leaching of rock it is convenient to consider the various techniques under two broad classifications, percolation or static leaching and agitated or suspension leaching; whilst the former is generally used for lower grade material than the latter, economics and local conditions can cause different techniques to be contemplated for the same grade of material creating an even less distinct classification. Percolation leaching includes methods such as in-situ leaching, dump and heap leaching and vat and thin-layer leaching, whilst agitated leaching encompasses those techniques where slurries are enclosed in a leach vessel such as an agitated tank or autoclave.

An important choice which has to be made when developing a treatment route is the lixiviant which is to be used. For copper ores, this is almost without exception an aqueous solution of sulphuric acid, with perhaps ferric sulphide to generate acid and as a powerful and efficient oxidising agent for many copper sulphide minerals. Alternative reagents such as ammoniacal solutions and hydrochloric acid are used occasionally for treating N_i - C_u mattes and have been proposed for the leaching of copper sulphide flotation concentrates. Despite these proposals though, sulphuric acid predominates in copper hydrometallurgy due to its low cost, its rapid attack on oxidised ores and the low consumption when sulphide minerals are leached because of regenerative side reactions.

2.5.1 Dump and Heap Leaching

The leaching of low-grade copper bearing waste has been practiced either by accident or through design for several hundred years and is one of the oldest metallurgical operations known. Originally waste material was removed from the mine site and disposed of at minimum cost and with disregard for subsequent leaching. This has led to haphazard development and the neutralization of an enormous resource of copper without maximum extraction of its potential. However recently increasing attention has been paid to more systematic development and within the last 10 - 15 years there has been greater study of those factors which could lead to greater understanding and improved design of the leaching process.

The production of large dumps of copper bearing waste material is largely associated with mining, and especially open-pit mining, of porphyry copper deposits. Important minerals in such deposits are chalcopyrite ($CuFeS_2$), chalcocite (Cu_2S) and pyrite (FeS_2); the iron components of the mineralization are of particular importance for the production of ferric sulphate which contributes to the leaching of the copper sulphides and also for the regeneration of sulphuric acid.

Despite numerous investigations, these having been thoroughly reviewed by Murr (258), the mechanisms of the process are not completely understood, but it is known that autotrophic bacteria accelerate the leaching reactions to some extent. The two common bacteria are Thiobacillus Ferro-oxidans, which lives on and catalyses the oxidation of ferrous ions to ferric, and Thiobacillus Thio-oxidans, which appears to contribute to sulphide oxidation, though it is not clear how. The main reactions involved in leaching sulphide minerals are thought to be:-

a) Introduction of ferrous ions into solution by the action of sulphuric acid and oxygen on iron sulphide minerals and/or oxidation of chalcocite to covellite:

$$CuFeS_2 + 4O_2 \rightarrow CuSO_4 + FeSO_4$$

$$H_2O + 2FeS_2 + 7\tfrac{1}{2}\ O_2 \rightarrow Fe_2(SO_4)_3 + H_2SO_4$$

$$Cu_2S + H_2SO_4 + \tfrac{1}{2}O_2 \rightarrow CuS + CuSO_4 + H_2O$$

b) Bacterial oxidation of ferrous ions and covellite

$$2\ FeSO_4 + H_2SO_4 + \tfrac{1}{2}O_2 \xrightarrow{bact} Fe_2(SO_4)_3 + H_2O$$

$$CuS + 2O_2 \xrightarrow{bact} CuSO_4$$

c) Ferric sulphate leaching of sulphide minerals

$$Fe_2(SO_4)_3 + Cu_2S + 2O_2 \rightarrow 2FeSO_4 + 2CuSO_4$$

$$2Fe_2(SO_4)_3 + CuFeS_2 + 3O_2 + 2H_2O \rightarrow 5FeSO_4 + CuSO_4 + 2H_2SO_4$$

$$Fe_2(SO_4)_3 + FeS_2 + 3O_2 + 2H_2O \rightarrow 3FeSO_4 + 2H_2SO_4$$

The cyclic nature of the reactions in (b) and (c) suggests the generation and neutralization of a great deal of acid within the dump, so much so that, generally, it is unnecessary to add make-up acid to the system to maintain leaching. Similarly the amount of iron that is solubilized and precipitated is such that the replacement of cementation by solvent extraction and electrowinning for copper recovery is unlikely to have any significant effect on the iron balance in the process.

Heap leaching is in most respects the same as dump leaching; however heaps are generally of smaller size than dumps and are deliberately constructed with leaching in mind. The method is used to treat oxide ore deposits which are often extracted during the surface stripping of prophyry deposits, and although sulphide minerals may be present, recoveries of copper from these are low due to the fairly short leaching times (usually 100-180 days). However it should be noted that the process and its analysis are the same when dealing with oxide dumps which have built up around many mines over the years. The oxide minerals are the degradation products of what were originally sulphide deposits; table 2.1 lists the most important.

Table 2.1 Economically important copper oxide ores

Type	Mineral	Formula
Carbonates	Azurite	$2CuCO_3.Cu(OH)_2$
	Malachite	$CuCO_3.Cu(OH)_2$
Silicates	Chrysocolla	$CuSiO_3.2H_2O$
Sulphites	Antlerite	$3CuO.SO_3.2H_2O$

The reaction mechanisms involved in dissolution using sulphuric acid are, it is believed, much simpler than for sulphide leaching. They can be illustrated by the equilibrium equation for chrysocolla:-

$$CuSiO_3 \cdot 2H_2O + 2H^+ \rightleftarrows Cu^{2+} + SiO_2 \cdot nH_2O + (3-n)\ H_2O$$

As can be seen there is no acid regeneration and so consumption is often a critical factor, especially when the copper is associated with carbonaceous, acid consuming gangue. Another important consideration is the more rapid reaction kinetics involved.

In contrast to the number of papers discussing the theoretical aspects of dump leaching, there have been very few reports on the evaluation of ores for leaching and on the practical aspects of the operation of dumps. Bruynesteyn and Duncan (259, 260) have described the methods used at B.C. Research, one of the experienced test facilities in this area, and it would appear, from discussions and implications and vague descriptions in other papers, reasonable to assume that their straightforward, 2 stage approach is common to most evaluations.

Once mineralogical and analytical studies have indicated the presence of economic quantities of minerals, a small representative sample is ground and leached under optimum conditions for 1-2 weeks in a flask or similar vessel. These tests should show whether there is any reasonable extraction and/or excessive acid consumption. Ores which pass this preliminary testing are then studied in column leaches. The amount of sample required for these tests is generally at least 1 tonne and although the type of column is not too important, its diameter should be large enough to swamp any wall-effects. Leaching is then carried out under virtually pilot-plant conditions, with some type of extraction system to reduce the copper content of the pregnant leach liquor. These tests tend to be of fairly long duration, often approaching 6 months in length for a sulphide material, due to the slow nature of the process. During the tests the ore and the leach solution would be sampled at regular intervals to monitor the state of the reactions and produce a time vs recovery curve. It would be upon a necessarily subjective analysis of the data accruing from these tests that recommendations for treating the material and forecasts of performance are made.

The length of these tests and the amount of sample required potentially can cause a major problem when they need to be incorporated into the normal project evaluation program. The difficulty is that leaching processes are more complex in terms of their mineralogy, chemistry and fluid flow and take 1,000 to 10,000 times longer than

the other unit operations that are normally encountered (261). Bartlett (262) and Roach and Prosser (263-266) have attempted to produce models to enable leaching behaviour to be predicted without column leach tests. Such predictions would be based entirely on basic measurements and analyses made on an ore in a laboratory, together with known chemical and physical parameters. Roach and Prosser's model is potentially difficult to apply to real materials due to an idealized model of pore structures; it has only been tested on artificial materials though the predictions of extraction were reasonable when compared to the actual performance. However both of these approaches require considerable further development before they are of practical and commercial value and it is therefore still necessary to carry out column tests to study certain variables, a number of samples and reasonable testing period being required.

It is therefore necessary to develop scale-up relationships or models to enhance the design and interpretation of experimental testwork and transform laboratory data into predicted field results (267). Whilst it would appear that scale-up from very large columns is satisfactory despite differences in prevailing temperatures (268) and that there is the promise of evidence to show that the use of data from laboratory column tests is valid (269), the scaling of static leaching is very complex due to the number of parameters and their interactions, and requires far more than a consideration of simple geometric parameterization. There are serious problems involved in obtaining satisfactory data with which to study this problem and so more detailed studies need to be carried out before there can be any confidence in the design of suitable experiments and the transformation of the resulting data into predictions of operating conditions (258, 267, 269).

The first analytical model of dump leaching was developed by Taylor and Whelan (270). However this required a knowledge of the past history of the dump and was unable to predict the effect of specific parameters as all factors were combined into two lumped constants. Hence, the model is not only restrictive in form but is also valueless in scale-up work.

A second model described by Harris (271) suffered from a similar limitation in that a difficult to determine size parameter was used

and there is also no simple means of simulating any change in this. The model was based on the hypothesis that diffusion of oxygen through the heap was rate controlling and that availability of oxygen and surface area for leaching are the most critical factors.

Since these early modelling efforts, research has been oriented at developing models to predict the performance of a heap from laboratory tests. Whilst these models have important differences, nevertheless they all start with essentially the same view of the particle leaching process. It is assumed that those mineral particles near the surface of a rock fragment tend to react first and that the reaction zone moves inward towards the centre; for mathematical simplicity, equations are derived using the assumption of spherical fragments with a uniform mineral distribution and a uniform progression of the reaction. It is also assumed that each fragment is continuously and completely surrounded by the leaching agent. Whilst these assumptions are perhaps acceptable for reasons of simplicity, their validity is particularly questionable.

There are three main groups of model, all of them based on the concept of a shrinking core of unreacted material, shown schematically in Fig. 2.13. The overall reaction process can be divided into three steps; external mass transfer, diffusion through the product layer and the chemical reaction at the interface between unreacted and reacted zones. Any of these steps can be rate controlling and equations applying to each one derived before combination into an integrated overall rate expression for the system. Treatment of such systems is quite common in process analysis and is of fairly standard form (272-274) being used for gas-solid as well as liquid-solid reactions. The three groups differ only in the assumptions made to solve the system describing equations, these being largely concerning the degree of control by a particular process step; fig. 2.14 (267) shows the basic differences between the models. Bartlett (262) assumes that the chemical reaction rate is slow resulting in a large reaction zone (fig. 2.14(a)) containing mineral grains at various stages of leaching. The model developed by Braun et al (275) assumes that all of the grains within the reaction zone are at the same stage of leaching, implying a narrower reaction zone and therefore faster reaction kinetics (fig. 2.14(b)). Roman et al (276) assumed instantaneous reaction kinetics and therefore a diffusion controlled process (fig. 2.14(c)).

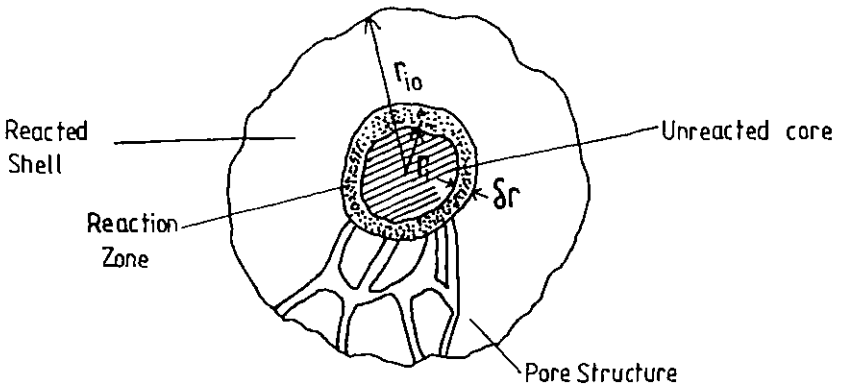


Figure 2.13 Ore particle of radius r showing reaction zone and unreacted core

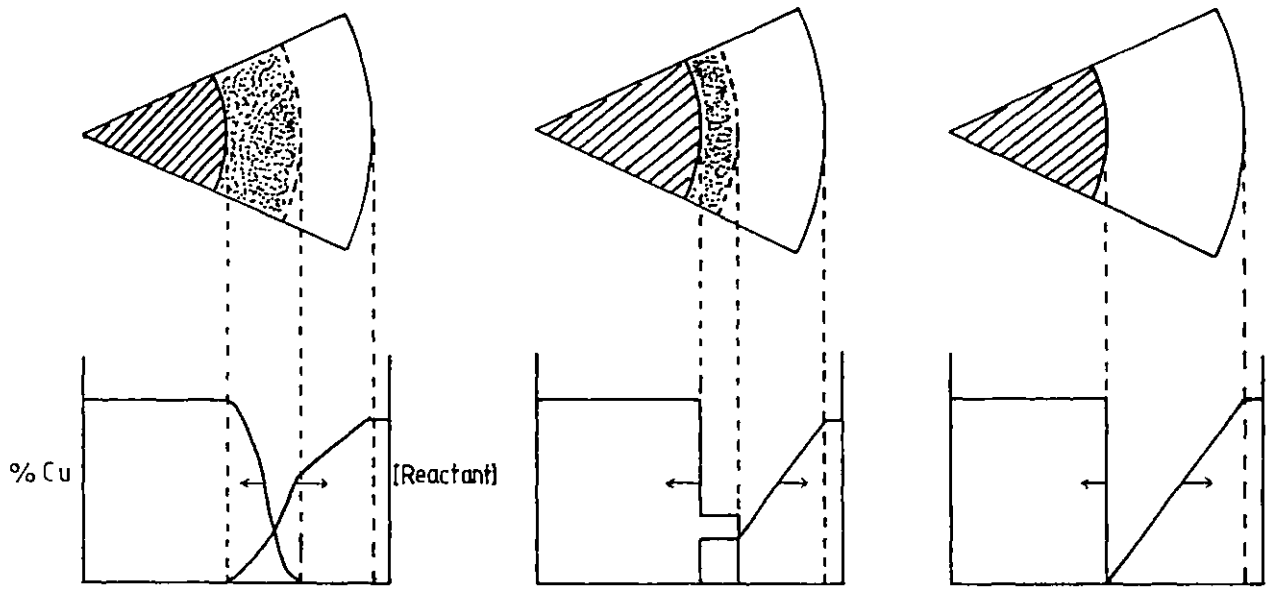


Figure 2.14 Three groups of models proposed for particle leaching (267)

Bartlett's (262) approach was perhaps the most rigorous, coupling a diffusion rate equation with an equation describing chemical rate of reaction to produce a continuity equation describing copper extraction as dependent on both time and radial position. It was intended that it would only be necessary to input physical and mineralogical parameters and chemical reaction rate data which were obtainable without resort to column testing and indeed reasonable correlation of column leaching data was achieved using this model (277, 278). However it was assumed that the overall extraction rate was controlled by processes within the rock fragment and as such it was difficult to couple with dump or solution parameters; also the computational solution of the model was excessively complex. With regard to the measurement of model parameters, some, such as tortuosity, were difficult to measure and the correlations mentioned above were achieved using data from the column tests themselves, thus negating the advantages of the model's form.

Madsen and Wadsworth (279) modified Bartlett's model, assuming a quasi-steady state to simplify the computational algorithm and utilizing curve fitting to supplement the measured parameters; using this approach it was possible to predict the results of one test based on data from another test on the same material, quite good correlations being obtained for a number of materials. Despite including an algorithm for calculating the ferric ion concentration in the bulk solution, this model is still insensitive to macroscopic dump and solution parameters. In the model, intrinsic mineral reaction rates were used in a steady state approximation of the continuity equation to determine the pore diffusion of ferric ions to the sulphide mineral surface, thereby enabling copper extraction to be calculated; the continuity equation for ferric ions in a rock particle is:-

$$\varepsilon \frac{\partial\{Fe^{3+}\}}{\partial t} = \Sigma R_k + D_e \nabla^2 \{Fe^{3+}\}$$

where ε = ore porosity

D_e = effective ferric ion diffusion coefficient

∇ = standard differential operator

and $\Sigma R_k = D_e \left(\frac{d^2C}{dr^2} + \frac{2dC}{rdr} \right)$

= rate of consumption of ferric ions

The relationship between ferric ion consumption rate and mineral leaching rates is given by:-

$$\sum_k R_{jk} = \sum_k \left[\frac{\sigma_k \ell_r G}{MW_{cu}}\right] \left[\left(\frac{d\alpha}{dt}\right)_{jk}\right]$$

where $\sum_k R_{jk}$ = rate of Fe^{3+} consumption at position j by mineral k

$\left(\frac{d\alpha}{dt}\right)_{jk}$ = function of $\{Fe^{3+}\}$ at j and fraction reacted of mineral k

σ_k = stoichiometry factor $\frac{(\text{moles of } Fe^{3+}}{(\text{moles of } Cu2+ \text{ extracted from mineral k}}$

ℓ_r = ore density

G = ore grade

MW_{cu} = mol. wt. of copper

The leaching kinetics of individual ferric-ion consuming minerals are necessary to describe the rate of ferric ion consumption term (258).

Braun, Lewis and Wadsworth (275) developed a model initially for in-situ leaching which has been used by a number of workers for dump leaching situations. It was assumed that circulation of solution around the particles is sufficient to maintain reactant concentration at the particle surface at the concentration in the bulk solution so that bulk solution transport is not a controlling factor. A narrow reaction zone is also assumed, within which there is uniform solution concentration and extent of copper extraction. The equation derived for the leaching rate, based on shrinking core models developed for high temperature gas-solid reactions, was:-

$$\left(\frac{dn}{dt}\right)_i = \frac{-4\Pi r_i^2}{\phi_{io}} \cdot C_b \cdot \frac{1}{\frac{1}{G\beta} + \left(\frac{\sigma}{Df}\right)\left(\frac{r_i}{r_{io}}\right)(r_{io} - r_i)}$$

where $\beta = \frac{3\ \ell_r\ \delta\ k}{r_p\ \ell_p}$

$G = \frac{n_p\ A_p\ r_p\ \ell_p}{3\ \ell_r}$

and:-

r_i = radius unreacted core at time t δ = thickness of reaction zone

$\emptyset_{io}$ = geometry factor

C_b = molar reagent concentration in bulk solution

σ = stoichiometry factor

D = diffusion coefficient

f = porosity and tortuosity factor

r_{io} = original particle radius

ℓ_r = rock density

k = chemical reaction rate constant

r_p = average radius of mineral particles

ℓ_p = average density of mineral particles

n_p = $\frac{\text{number of mineral particles}}{\text{unit volume of rock}}$

A_p = average surface area of mineral particles

As can be seen the formulation includes fairly arbitrary and difficult to measure factors to allow for particle geometry reaction zone thickness, porosity and tortuosity; indeed, whilst it was possible to use the model successfully to describe chalcopyrite leaching, it was found necessary to allow the geometry factor to change with time as it was difficult to determine a constant value to fit the data. It was also necessary to introduce empirical parameters to allow for early rate enhancement, due to the mineral richness at the outside of the particles, and late rate enhancement, possibly due to the generation of cracks and fissures in the rock.

Madsen, Wadsworth and Groves (280) used this model to describe the dilute sulphuric acid leaching of sulphide waste, utilizing the results from one test to satisfactorily predict the results from a second and from 1 particle size to another. They found the use of rate enhancement factors unnecessary if the diffusion coefficient and reaction rate constants were allowed to change with time; initially the process was reaction controlled but with time, diffusion control increased. Madsen et al (280) concluded that Braun et al's model (275) was useful in this modified form but that problems occurred due to the particle shape parameter and if solution flow was not ideal. Using 5 tonne samples of 3 different sulphide ores, Meixner (281) found that this form of the original model described leaching behaviour quite well and that once model parameters had been determined reliable extrapolation was allowed.

Two of the deficiencies of this group of models are their inability to include dump parameters and the lack of testing on a large scale leach dump. However Cathles et al (282-285) have used Braun et al's

model concepts, but extended them to include an oxygen balance, a heat balance and to allow for air convection throughout the dump. Initially a one dimensional model was derived and applied to data obtained from a small test dump, but this has been updated into a two dimensional version which was calibrated against large scale leach column experiments and the known leaching histories of five industrial dumps; the success of this model in these tests was considered satisfactory, and on the basis of these tests the Kennecott Copper Company used the model to derive simpler, empirical models (286) with which dumps could be designed and operating performance evaluated. In this respect the only deficiency in the development programme was the lack of calibration against laboratory columns, although earlier work (268) on the large columns had predicted that smaller column data was adequate for scale-up if temperature dependency was not a major factor; the Cathles et al model however indicated how important thermal effects could be in an operating dump.

Bacterial effects were introduced into the model by assuming that there is a temperature, T_{sick}, above which bacterial conversion of ferrous to ferric ion begins to decline, and a temperature, T_{kill}, at which bacterial catalysis ceases. It was then assumed that sulphide oxidation kinetics lose temperature dependence at T_{sick} and that the leach rate could be extrapolated linearly from its value at T_{sick} to zero at T_{kill}. Consequently if bacteria are not functioning, the model assumes sulphide oxidation to cease; this would appear to be born out in practice (258).

The rate of copper leaching is given by:-

$$R_{cu} = \ell_R (1 - \emptyset)(G_s \frac{dX_s}{dt} + G_{NS} \frac{dX_{NS}}{dt})$$

The rate of oxygen consumption is given by:-

$$R_{o2} = \ell_R (1 - \emptyset) G_S \frac{dX_S}{dt} (1.75 + 1.91 FPY)$$

The rate of heat generation is given by:-

$$R_A = \ell_R (1 - \emptyset) G_S \frac{dX_S}{dt} (2.89 + 5.41 FPY)$$

where

$$\frac{dX_S}{dt} = \frac{- 3X_s^{2/3}}{6 \tau_{Ds} X_s^{1/3} (1 - X_s^{1/3}) + \tau_{cs}}$$

$$\frac{dX_{NS}}{dt} = \frac{- 3X_{NS}^{2/3}}{6 \tau_{DNS} X_{NS}^{1/3} (1 - X_{NS}^{1/3}) + \tau_{CNS}}$$

$$\tau_{cs} = \frac{K_R}{K_{ox} A^R_{sulf} \delta \{O_x\}}$$

$$\tau_{Ds} = \frac{T_R' a^2 K}{6\{O_x\} D_{ox} \phi_n'}$$

No guidance is given on the calculation of τ_{DNS} or τ_{CNS}, so it must be presumed that these are analagous to τ_{Ds} and τ_{cs}

The temperature dependence of τ is given by:-

$$\tau(T^oC) = \tau(0^oC) \exp \frac{1000.E^*}{R} \frac{T}{273(273+T)}$$

where the model parameters are:-

ℓ_R = rock density

ϕ = inter block dump porosity

t = time

FPY = mole ratio pyrite/sulphide Cu: X_s + X_{NS} = fraction of initial Cu type remaining in leach unit

τ_D = time to completely leach a particle under diffusion control

τ_c = time to completely leach a particle under unreacted core surface area

K = oxidant required to leach unit volume of rock

K_{OX} = rate constant for ferric oxidation of pyrite

a = waste particle radius

A^R_{sulf} = surface area of sulphide mineralization per unit volume

δ = reaction zone thickness ($\simeq$ a/10)

$\{O_x\}$ = concentration of ferric ions

T_R' = tortuosity of diffusion channels

D_{ox} = diffusion coefficient of ferric ions

ϕ_n' = porosity

E* = activation energy

X_s = fraction of sulphide Cu left

X_{NS} = fraction of non-sulphide Cu left

G_s + G_{NS} = respective Cu grades

No derivation is given for the above equations and their solution is a "fairly large undertaking" if the computer program were to be developed from scratch (287). Other than the use of Braun's model to describe the leaching reaction and the slightly arbitrary handling of bacterial action, the most questionable assumption inherent in this model is that the dump is composed of particles of a single size; this assumption is justified by the authors on the grounds that small ore particles tend to clump together and leach as if they were a larger aggregate. The only other potential problems are likely to arise from the prediction

of certain parameters, such as porosity and tortuosity, for an operating dump. Other than in the prediction of industrial performance from laboratory tests, this model appears to be of considerable potential and should provide the basis for further developments.

The third group of models is based on a model developed by Roman et al (288) and the assumption that there is a narrow, very sharply defined reaction zone when copper oxide minerals are leached; inherent in the model is the assumption that as the chemical reaction rate and film boundary diffusion rates are so high, the diffusion rate through the particle is rate controlling. The model formulation was verified through further testwork and application by Shafer et al (289). The form of the model, equivalent to the diffusion controlled part of Braun et al's model, is (289):-

$$1 - \frac{2}{3}\alpha_{ji} - (1 - \alpha_{ji})^{2/3} = \frac{2\ Deff\ M\ C_H\ t_i}{\ell_r\ \sigma\ G_j\ r_j^2}$$

where α_{ji} = predicted fraction of ore, size r_j, leached at time t_i

Deff = effective diffusion coefficient (combining porosity, tortuosity and particle shape factors)

M = molecular weight of ore constituent being leached

C_H = concentration of lixiviant

ℓ_r = ore bulk density

σ = molar acid consumption

G_j = grade of ore of particle size r_j

r_j = radius of particle

One complexity in the use of this model is that there is a consumption of acid in the leach reaction and, unlike sulphide leaching, it is unlikely that this will be regenerated through bacterial action. This reagent depletion is often increased by the presence of acid consuming rock and is especially critical near the bottom of the dump where the acid may be completely consumed if the dump is too high or feed concentration is inadequate. The model therefore assumes that the change in lixiviant concentration is directly proportional to the change in copper concentration. Unfortunately the proportionality constant can also be a variable being affected by particle size, acid concentration and the extent of reaction.

Another problem in the application of this model is that the model equations must be solved numerically. To overcome this, Roman (288)

developed a numerical analysis technique which involves treating column leaches as a series of finite differential volumes; each unit volume is placed on top of another in order to simulate a leach column and a number of columns together to make up a heap.

Although not a particularly easy model to use, this has proven to be fairly reliable and requires only a fairly short and small column test to provide the necessary parameters; once these have been determined, the model is quite flexible in its ability to assess the effect of changing variables as is witnessed by the economic and sensitivity analyses published based on its use (289-291). It should be noted though that model predictions have not been tested on a full scale heap and that there must be some difficulty in the prediction of parameters such as bulk density for a large heap.

Due to the possible inaccuracies inherent in Roman's allowance for acid consumption and the difficult numerical methods used to solve the model, Averill (292) and then Chae and Wadsworth (293) developed the original concept and incorporated continuity equations to describe the variation of reagent concentration with depth. However this approach has only been tested in the laboratory, but more importantly is only formulated and tested for small and single size ranges.

Summary and Discussion

Whilst it would appear that the development of models to date is of value, that the shrinking core concept is valid, and that the models have some predictive capability, it is equally apparent that they have many important limitations, have not been adquately tested, and therefore should be considered to give only a best possible view of the extraction profile.

The models are generically similar, though there are many differences in detail, these largely evolving from the amount of rate control given to the chemical reaction and reactant diffusion processes. To date each model has been tested on few different materials and it would seem reasonable to conclude that whilst each model may be suitable for a specific application, no model developed so far is suitable to be generally applicable.

It is notable that a number of factors usually accepted to be of importance in real systems have hardly ever been incorporated into models so far developed. One of the most important factors ignored is

the flow of solution; whilst Jacobsen (294) studied the flow through a dump, it is generally assumed in the models described that the solution flows vertically through the dump in a plug flow fashion and that all particles have an adequate degree of contact with the lixiviant They therefore ignore channelling, often caused by agglomeration of fine particles, precipitation of salts due to high pH and the effect of any clay minerals present. Other factors which are often ignored or are treated relatively arbitrarily include the effects of temperature and of bacterial action.

Whilst the general theoretical basis of the leaching models, the shrinking unreacted core, appears to be valid, the theoretical significance of some of the factors used to correlate the models with experimental data would seem to be questionable; this is highlighted in Braun et al's work where it was found necessary to adjust a factor, nominally allowing for particle geometry effects, as leaching progressed, and to introduce early and late rate enhancement factors. Often these factors, whilst dimensionally descriptive, cannot be measured directly but are determined from the results of a completed leach test and then used in the prediction of other test results.

These inadequacies lead on to the deficiencies associated with parameter determination and whether and how results of small laboratory column tests can be used in the prediction of industrial scale leaching performance. Considerably more work is required on modelling, scaling and the verification or calibration of models in order to gain more insight and confidence in the application of such techniques(258). Whilst shrinking core models have demonstrated considerable success on a laboratory scale or small pilot scale, but have rarely been tested directly in an actual industrial dump or on a scale which would begin to approach an operating dump. At present the design of static leaching processes and/or the determination of model parameters still requires testwork on a number of samples and lasting for a considerable period, especially with respect to the design of other processes. Not only is development necessary in verifying the suitability of small leach column testing, but also in reducing the amount of parameter determination; it is to be hoped that the approaches of Bartlett and Roach and Prosser will lead to improvements in this area.

With the present state of the art it would seem that Chae and

Wadsworth's model has the best potential for predicting oxide mineral leaching, whilst the performance of a mixed sulphide dump might best be modelled by a combination of Madsen and Wadsworth's description of the leach reaction and Cathles et al's approach to dump modelling. However these models and any others likely to be developed along similar lines, are inappropriate in form for a steady state process simulation of the type proposed in this project. The numerical methods required for their solution are excessively long to be included in a simulation. Also static leaching is a dynamic process and the models developed reflect this in that they only predict copper recovery at a particular time. Perhaps the best approach would be similar to that used by Kennecott and to develop a suitable model for the particular ore and situation being studied; once the model parameters had been determined from testwork, the model could then be solved for a planned number of situations and the results of these simulations used to predict the parameters of a generally applicable, empirical model which could be incorporated into a process simulation. Using this approach, the duration of the leach would be a design variable and solution concentration and recovery would perhaps be averaged over this life using the assumption that a number of such leaches were in operation to ensure a steady feed supply to the copper recovery sections. It is unlikely that a series of such models arranged in counter-current flow fashion would also simulate vat leaching plants, though, as almost no work on the design or modelling of these installations has been published, this is somewhat hypothetical and may be inappropriate.

2.5.2 Agitated Leaching

Agitated or suspension leaching techniques are used when the feed material is of high enough grade and reacts quickly enough to justify the extra costs involved; although several leach processes have been proposed and pilot plant tested for sulphide flotation concentrates, generally only oxide minerals are treated using this method due to the slow reaction kinetics of sulphide minerals and the short contact time afforded in an agitation tank. For oxide minerals leached by sulphuric acid the chemical reactions are the same in suspension leaching as in heap leaching. Generally the process is a continuous one and is carried out in a series of 3-5 tanks arranged in either a co-current or counter-current fashion with agitation being provided by air or mechanical means (257).

Little has been published on the testwork involved with agitation leach system design, though it would appear from discussions (295) that pilot plant testing is only used if the material proves to be problematical. However the scale-up of agitated tanks is a complex matter with a number of empirical expressions published but in general disagreement; even if strict geometric similarity is used, it is necessary to resort to some fairly dubious relationships to allow scale-up of the resulting data (296, 297).

The general objective of leaching experiments is to obtain an estimate of reaction kinetics over a certain design region of the process variables; this information is then combined with cost data to estimate the most economic operating conditions. From those methods published (298, 299) it would appear that these kinetic relationships and the general approach to design are largely empirical in nature.

Considerable research effort has been devoted to the formulation of generalized rate expressions such as the shrinking core model (300, 301) and the grain model (302), in an attempt to quantify the reaction kinetics and mechanisms of leaching systems. Unfortunately, these findings have for the most part been limited to homogeneous, monosize, sulphide particles in batch reactors, and relatively little has been accomplished on the modelling of multistage, continuous leaching systems. There seem to have been only two studies on the leaching of oxide minerals, Hsu and Murr (303) extending the experimental range of Pohlman and Olson (304) for the leaching of copper from chrysocolla using sulphuric acid; a simple chemically controlled model based on shrinking core concepts was developed but the application of this has been limited due to the restrictive assumptions of no diffusion control and homogeneous particles with only chrysocolla reacting.

As has been discussed by Sepulveda and Herbst (305), the task of modelling multiparticle leaching systems is far more complex than that of homogeneous ones. Added complexity arises from a number of sources, including the multiplicity of possible reaction rate controlling steps associated with heterogeneous reactions, the natural occurrence of a broad distribution of particle properties, interactions between particles, the effect of solution chemistry and electrochemical phenomena, each of these sources having a potential effect both directly and through interactions with others (306). Perhaps the most significant

omission is that of the reaction of minerals other than the main copper bearing mineral, either as a contributor to the production of copper or in terms of other effects such as the alteration of solution chemistry, especially in terms of acid consumption. This omission has been made by both Bartlett (307, 308) and Sepulveda and Herbst (305) in what appears to be the only two phenomenological modelling studies of continuous, agitated leaching systems.

Bartlett (307, 308) used a classical population balance approach combined with shrinking core model concepts to produce recovery - reaction time graphs for a number of reaction type - residence time combinations; however highly restrictive conditions, such as Gates-Gaudin-Schumann feed size distribution, only one rate controlling step at a time, and failing to account for lixiviant depletion, limit the application of these graphs even further. Sepulveda and Herbst (305) also used population balance concepts combined with shrinking core models to develop a modelling approach to continuous leaching systems. Apart from the limitation inherent in only considering one mineral reaction, their modelling is restricted by the complexity of equations derived and the difficulty in solving these numerically for a particular system; however their techniques should be developed into simpler, more applicable models (305) and, if these can be combined with the modelling of more than one mineral reaction, could lead to a very useful process design method.

Summary and Conclusions

For the present it would appear that the phenomenological modelling of leach kinetics is too complex, or insufficiently developed, to provide a reasonable simulation of agitated leaching. Perhaps the best approach is to combine residence time distribution models, which are likely to be incorporated into future process models, with an empirical modelling of the leaching kinetics. Whilst there are many ways in which the kinetics modelling could be handled, Christie and Welch (309) have proposed one which is soundly based on statistical theory and provides economical use of experimental resources. Parameters for a second order statistical model can be determined from a designed experimental programme, the important process variables for a particular system being decided upon by the design engineer; separate models for copper extraction and acid consumption could be easily derived from such

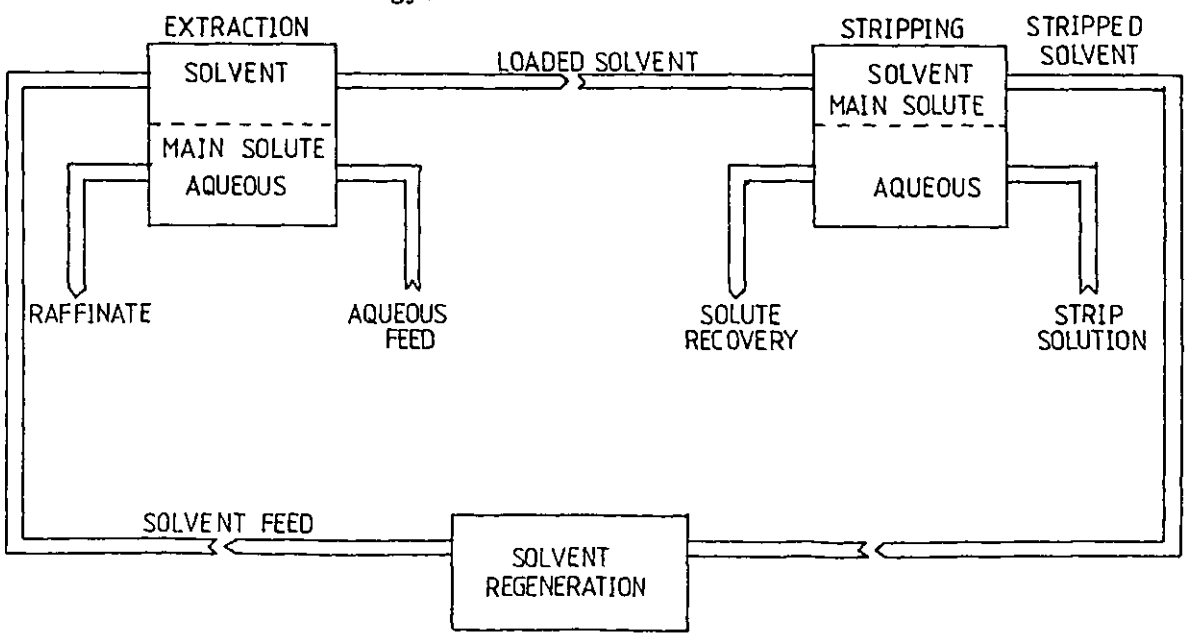
a programme and combined with a flexible model of tank residence time to produce a simple but robust model of the leaching process.

2.6 Liquid - liquid extraction

2.6.1 Introduction

Liquid-liquid extraction (more commonly, though perhaps incorrectly (310), known as solvent extraction) is now an established metallurgical unit process, frequently included in the process flowsheets of hydrometallurgical plants. From a relatively unknown process twenty years ago, restricted in application to the nuclear field, some less common metals and the analytical laboratory, it has found many large scale applications in the petrochemical, pharmaceutical and metallurgical industries. In hydrometallurgy, liquid-liquid extraction has become one of the most important processes for separation, purification and concentration of metal ions and inorganic salts, though due to the high cost of solvents, industrial applications are limited at present to the more valuable metal products (311-313).

The method is used for separating the components of a homogeneous liquid mixture by distributing them between the original solvent and a second liquid which is either immiscible or only partially miscible with it; this is generally achieved using an organic liquid as the extracting solvent. A second, and equally important, result of the extraction can be to increase the concentration of the metal ions in solution. Solvent extraction is applied in hydrometallurgy to separate the valuable metal from any impurities which would contaminate the final product and to increase metal concentration in the feed to the precipitation process. Fig. 2.15 shows schematically the process and its associated terminology.



In metallurgical applications the solvent is rarely a single liquid but rather consists of a diluent in which is dissolved an organic compound called the extractant; this may have acidic properties (extracts cations), or basic properties (extracts anions) or may be a compound that solvates neutral species more strongly than water. Extraction is achieved by interaction of the extractant with the metal species of interest in the aqueous phase to form a complex or solvate soluble in the organic phase. Selectivity is achieved through the nature of the extractant-metal species interaction and the prevailing chemistry of the aqueous phase.

Nowadays a very large number of extractants are available for commercial use in hydrometallurgy. More than 40 proprietary reagents of all types are available, whilst at least a dozen are in everyday use.

2.6.2 Liquid - liquid Extraction Plant Design

Hanson (314) has recently outlined 3 broad facets to be considered when approaching any potential application of the technique:-

1) Development and design of the process; this involves selection or development of a solvent system, demanding consideration of both equilibrium and kinetic data.
2) Selection of the optimum cont actor for the system, bearing in mind the degree of separation required, the rate of interphase transfer and the need for stable operating characteristics.
3) Interaction of the solvent extraction unit with other parts of the process and with the environment.

These are, of course, subject to the final overall criterion:- the degree of economic success achieved.

From studies of the economics (315 -317) it can be deduced that the important areas are:-

1) number of stages required for efficient extraction
2) stage efficiency
3) flow capacities and phase ratios.

The capital costs depend largely on the size and number of contacting stages and the solvent inventory. The only other significant costs are operating and maintenance labour and the power required for pumping and mixing. The overall economics also rely on the efficiency of the process and the proportion of solvent lost to the aqueous streams.

Although it would be useful to have a definitive design procedure,

there is no such universal procedure known at the moment; work on large scale equipment has mainly been at an empirical level giving certain rule of thumb design methods, but those correlations are in the main restricted to the particular type of equipment used (318).

However from a broad survey of the literature a possible 3 stage design procedure would be:-

1) Define the aqueous chemistry of the system so that the appropriate class(es) of solvents can be selected and screen test these on particular feed solution.
2) Once promising systems have been identified, on the basis of satisfactory extraction, stripping and phase separation, isotherms for extraction and stripping are developed and the mixing kinetics and settling rates determined; these can then be used to calculate the number of theoretical stages required for particular flow rates and stage efficiencies and a preliminary cost estimate can then be produced.
3) Further systems development work to determine extraction and stripping rates, coalescence rates, entrainment levels etc. should be carried out, and it may be necessary for continuous testing on a laboratory scale to check the process chemistry and/or on a pilot scale to check the process engineering, before proceeding to final plant design and costing.

Ritcey and Ashbrook (313) give a good coverage of the techniques used in metallurgical applications with examples of actual plant studies.

However for some processes a considerable body of information is available which can give a sound basis for estimating process parameters in the absence of detailed experimental work. The recovery of copper and uranium from acidic solutions are examples of such processes. It must be remembered though that leach solution characteristics can be a crucial variable so that some confirmatory work should be carried out on samples of the actual solution.

Equipment Selection Once the chosen system has been evaluated, an extremely important consideration is the choice and design of contractor. This choice involves many factors, both chemical and economic (318-320) and is further complicated by the number of different cont actors that have been developed for liquid-liquid extraction; indeed the number of cont actors described in the literature is disproportionately large in comparison with the current industrial practices (310). Bailes et al (310),

Treybal (311), Hanson (312), Robbins (320) and Mumford (321) have reviewed these comprehensively.

Unfortunately data on the performance of individual cont actors published in the literature have been derived with particular systems. Attempts to compare the performance of cont actors is therefore difficult because differences caused by different systems cannot be clearly separated from those inherent in the different cont actors.

However, the choice of contactor for hydrometallurgical applications has consistently been a variation of mixer-settler and is likely to remain so until suitable quickly reacting vessels have been introduced (322). This is largely because mixer-settlers are particularly suitable for operations requiring a high flow capacity, long residence times and few stages; other advantages are their flexibility, reliability, ease of operation and maintenance, and that they can be scaled up with relative confidence. The main disadvantages of mixer-settlers are the large ground area requirements, the amount of piping needed and the large solvent inventory involved; these problems significantly affect the capital costs of a project and can be of crucial importance in an economic evaluation of the process.

In the sizing of mixer-settlers it is common practice to regard the mixer and settler sections as separate units and hence study the kinetics and coalescence separately. Although this may be useful during preliminary assessments, the interaction of the two units cannot be ignored in detailed design work, the effect of varying the degree of mixing on settling rates is particularly significant.

The system's efficiency for a certain mixer design is particularly difficult to calculate accurately as it is not only dependent on physico-chemical factors but also on the design and operation of the mixer itself. As can be seen from the reviews of available contacting equipment (e.g. 323) there is a large choice even when only considering mixer-settlers, and each mixer will have unique characteristics, making generalizations difficult and only approximate. Indeed a recent paper by Barnea (324) argues that even a qualitative understanding of the phenomena and mechanisms involved is still lacking and that the full characterization of a liquid-liquid contacting system is impossible. Similarly a review of kinetic studies of metals extraction by hydroxy-oximes (e.g. 325) highlights the differing interpretations of basic

rate data with respect to reaction order and the rate controlling step. For instance at a 1980 conference one paper (326) claimed mixed kinetics with both diffusion and chemical reaction rates controlling for this type of extraction whilst a second (327) claims only the formation of chelates to be controlling. It would appear that for design purposes any useful rate models or relationships must have been obtained with apparatus of similar proportions to the industrial-scale equipment to be used, an onerous restriction considering the number of designs to be studied. Similar restrictions apply to the empirical equations for copper extraction proposed by Slater et al (328) which relate stirring speed, residence time and phase ratio but are dependent on mixer type, aqueous feed composition and organic mix.

There is therefore good reason to agree with Barnea (324) that "liquid-liquid contacting has not yet completely crossed the boundary between art and science". It would appear from the literature (313) and discussions (e.g. 329) that detailed design is carried out using small scale equipment and replicating actual conditions as closely as possible. When less accuracy is required estimates based on experience tend to be made.

This is a similar situation to that appertaining in settler design. The relationship between small batch tests and continuous flow phase-separation has been discussed recently (330, 331) and the conclusion drawn was that settling rates in continuous flow systems cannot be predicted from batch settling data and so considerable research is required, bearing in mind that the settling process is also dependent on mixer characteristics. Hence settler size is estimated as a function of dispersion band thickness at a specific settling rate i.e. the volume of dispersion settling per unit area of settler per unit area; this function is normally quantified using the designer's experience.

Before a material balance of the process can be drawn up and its effectiveness measured a study of the process equilibria is necessary. However the winning of such data is a time consuming operation and so a considerable amount of work has been carried out in an attempt to be able to predict or, at least, correlate equilibrium data. For metals systems involving chemical interactions, models have been proposed based on these interactions, although this approach has not so far been noticeably more successful than empirical or semi-empirical

methods (314). Forrest and Hughes (332) have classified the types of models available for metal systems between chemically based models and totally empirical models, though the plant designer will normally use semi-empirical and empirical models to describe the system (329).

Mass balances

Once stage equilibria and efficiencies have been quantified it is possible to calculate a mass balance around the circuit and the number of stages required to achieve satisfactory extraction. The techniques used for this computation are analagous to the design methods developed by chemical engineers for vapour-liquid separations such as distillations; in this field the McCabe-Thiele method is most widely used for multi-stage counter-current extraction processes. Although a number of authors (333, 334) advocate the use of such diagrams for estimating the required number of stages for a particular separation, such simple calculations inadequately predict the quantitative effects of system parameters or operating conditions on process performance (335). Metal extraction equilibria are too complex to be represented by a single two-dimensional distribution curve for an entire cascade; especially as the solution composition and pH will vary from stage to stage.

More appropriate methods have been proposed by Robinson and Paynter (336), who presented a family of curves of organic metal plotted against aqueous metal concentrations at constant acid concentration, and Hughes et al (337) who suggest the use of a three-dimensional surface plot of concentrations. These have been largely superseded by computer techniques which, whilst essentially numerical evaluations of the graphical situations, are capable of greater speed, accuracy and flexibility; extra parameters which can affect equilibria, such as temperature and non-ideality, may be programmed in. Examples of such programs can be found in the atomic energy industry (338), rare earths extraction (339-340) and copper - LIX systems (317, 336, 342).

The only other important factors to be considered in the design of liquid-liquid extraction plants are those relating to the interaction of the plant with other sections of the process. Of these the most important in cost terms is the entrainment of organic solvent in the aqueous phase. If carried in the strip solution to an electrolytic plant, reagents can produce burn on the cathodes which down grades the quality.

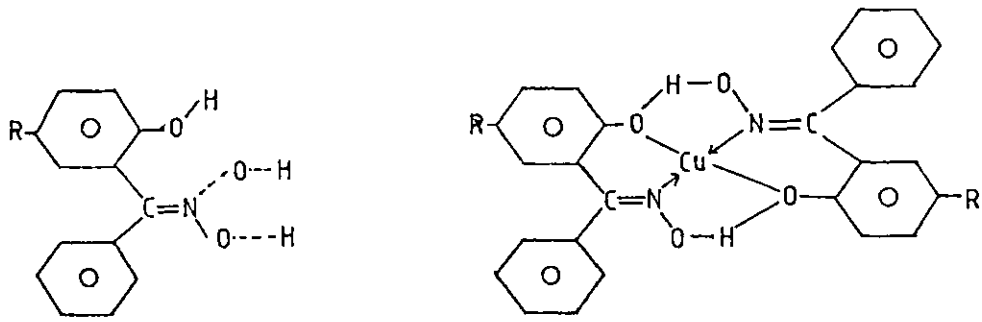
Similarly if organic is carried to a counter current decantation circuit it can absorb on to the solid particles and cause flotation. Aside from causing such operating problems, solvent make-up to allow for such losses constitutes one of the major costs in many extraction processes. However such losses are very difficult to quantify, and so, in the absence of experimentally determined losses, an assessment of the need for make-up is usually made based on experience. Coalescers or flotation machines are the most popular of solvent recovery, the choice depending on various operating and economic criteria.

In an analagous fashion the effects of upstream processes are difficult to measure; although solution impurities can affect the chemistry of extraction, the main problem is solids carried in the aqueous phase. Over a certain critical value these can cause crud growth resulting in further solvent losses. Generally, expensive solid-liquid separation stages have to be used to ensure a clear solution; these can include clarifying tanks and/or pressure filtration. Again, these effects are difficult to quantify and assessment must be made based on experience.

2.6.3 Copper liquid - liquid extraction systems

Introduction: Development of specific reagents for the extraction of copper from leach liquors came in the 1960's when hydrometallurgical processes were increasingly in demand and an alternative to cementation for recovering copper was sought after. Readily available reagents, such as organic acids, were generally unselective between copper and ferric ions and so intensive research was carried out by chemical manufacturers to develop a specific reagent for copper (343). Probably the best known of the resulting extractants are the LIX family produced by General Mills Inc. and the KELEX reagents from Ashland Chemical Co.; other reagents such as the P5100 range (ACORGA Ltd.) and SME 259 (Shell Chemical Ltd.) have also recently become available.

Although the properties and advantages of all these reagents are claimed to be many and varied, the active constituent in each one is an aromatic β-hydroxyoxime, such as shown in Fig. 2.16 for LIX 65N.



(a) (b)

Figure 2.16 LIX 65N (a) and its copper chelated species (b)

As the reagent structure can be generalized, so too can the stoichiometry of the overall reaction:-

$$Cu^{2+}_{(a)} + 2HL_{(o)} \rightleftharpoons CuL_{2(o)} + 2H^{+}_{(a)}$$

the subscripts (a) and (o) referring to the aqueous and organic phases respectively and HL representing the reagent molecule.

It is the reversibility of the reaction that lends itself to the application of these reagents commercially, as the mixing of strongly acidic solutions, such as waste electrolyte, with loaded organic phase returns the copper to the aqueous phase in a form suitable for winning the metal directly.

To date there are about 14 liquid-liquid extraction plants for copper recovery in operation or under construction. All but two of these use LIX 64N as the extractant, the exceptions being Ray Mines' new plant which uses P5100 and a small operation at Rustenburg where a carboxylic acid, Versatic 911, is used to separate copper from nickel. A summary of some of these plants is given in Table 2.2.

Of the other reagents KELEX 100 has often been evaluated as an alternative to LIX 64N, but has never reached commercial application despite promise with more concentrated leach liquors.

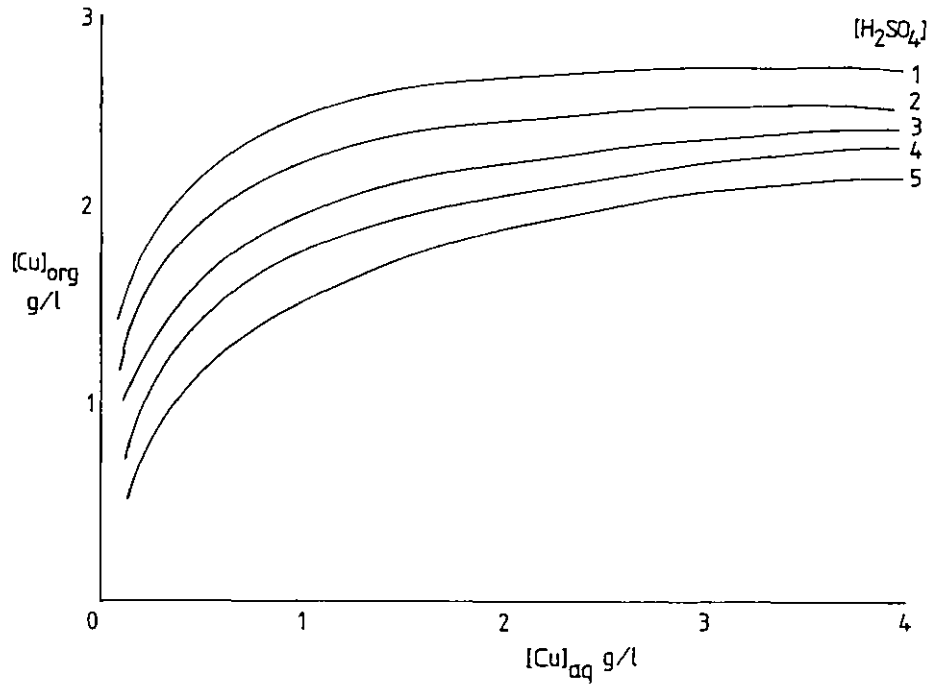
Modelling Equilibria. As suggested earlier, using a proven reagent such as LIX 64N it would be justifiable to reduce the solvent testing stage by using equilibrium isotherms developed from previous experimental work. Robinson and Paynter (336) produced a series of isotherms using batch laboratory-scale experiments and empirically derived equations from this data expressing the organic copper concentration as a function of aqueous copper, acid and LIX 64N concentrations. Forrest and Hughes (332, 344) have fitted this data to an equilibrium surface (337) and found an average error of only 4%.

However this data has been discredited by Whewell and Hughes (345) due to non-equilibration of the phases causing an anomalous "dip" in the isotherms at high copper and low acid concentrations.

Whewell and Hughes (345) then produced a fresh, comprehensive set of data which removed this problem and used this in the theoretical modelling of the system. However it was recognized that this data was potentially of use in fitting a new distribution surface and so regression analysis was carried out using the MINITAB program (88); it was found that there was insufficient data to provide an accurate model which included the effect of LIX concentration. However it was possible using a second order polynomial as suggested by Robinson and Paynter to fit surfaces for both 10% and 20% LIX concentrations with a correlation coefficient of 99.1% in both cases. The coefficients are given in Fig. 2.17(b) and 2.18(b) and plots derived from the models are shown in Fig.2.17(a) and 2.18(a). It was found that it was possible to fit one surface to the whole of the data unlike the work on the original data, though it was necessary to leave out the (Cu)aq. log {100(Cu)aq.} term due to auto-correlation. Although restricted in application, as is any empirical model, this type of model has been of aid in design work (345) and as such is of use in the proposed methodology.

Due to the complexity of the chemical reactions involved, deterministic models have been less successful for the copper - LIX 64N system. Forrest and Hughes (344) achieved reasonable accuracy (± 7% rel) but new parameter values were required for each constant acid isotherm, reducing the model's flexibility and usefulness. Hoh and Bautista's model (346) suffers from the same problem and is also restricted to dilute solutions due to the omission of aggregation in the organic phase; another drawback with their work is the assumption of 100% active oxime within the LIX sample (345).

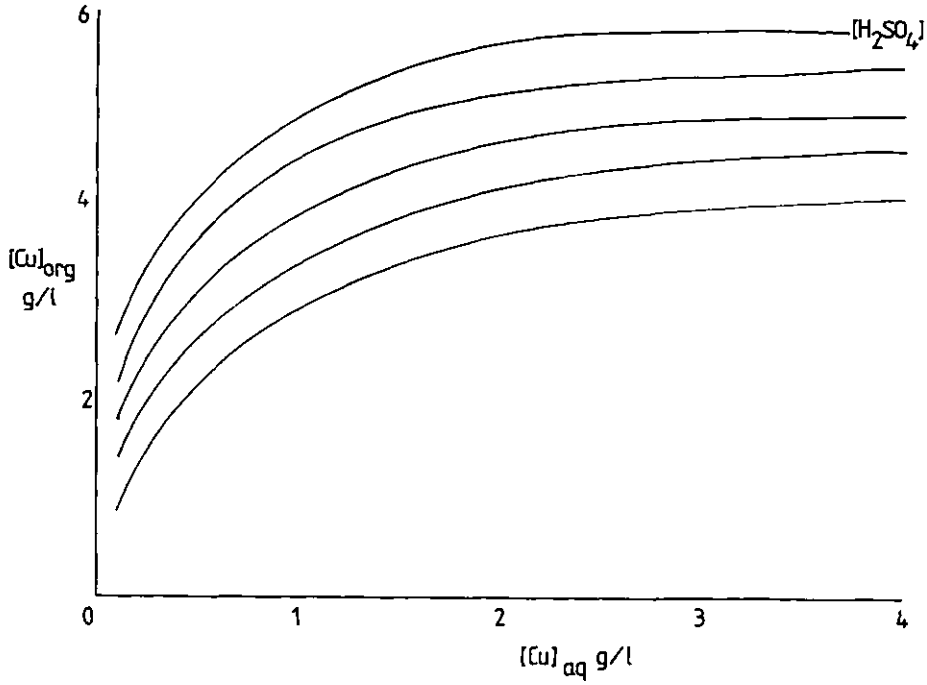
Perhaps the most successful phenomenological model of this system is that of Whewell and Hughes (345). This considers more sophisticated chemical forms, taking account of bisulphate formation and organic polymerization, an almost essential approach for the LIX reagents, though one which results in a series of fairly complex equations, the solution of which needs iterative techniques. Hence, although particularly accurate (7% ave. error), the model's application in a simulation program would result in excessive consumption of computer time.



$$[Cu]_{org} = 0.88 - 0.451.[Cu]_{aq} - 0.2951.[H_2SO_4] + 0.641.\log(100.[Cu]_{aq})$$
$$+ 0.0247([Cu]_{aq})^2 + 0.082([H_2SO_4])^2 + 0.257(\log(100.[Cu]_{aq}))^2$$
$$+ 0.0336.[Cu]_{aq}.[H_2SO_4] - 0.0052.[H_2SO_4].\log(100.[Cu]_{aq})$$

$R^2 = 99.3\%$ St.dev. = 0.0812

FIG. 2.17: 10% LIX 64N Equilibria model



$$[Cu]_{org} = 1.279 - 0.595.[Cu]_{aq} - 0.3537.[H_2SO_4] + 1.354.\log(100.[Cu]_{aq})$$
$$+ 0.0034([Cu]_{aq})^2 + 0.0163[H_2SO_4])^2 + 0.546(\log(100.[Cu]_{aq}))^2$$
$$+ 0.0629.[Cu]_{aq}[H_2SO_4] - 0.1289.[H_2SO_4].\log(100\,[Cu]_{aq})$$

$R^2 = 99.3\%$ St.dev. = 0.164

FIG. 2.18: 20 % LIX 64N Equilibria Model

The modelling of the copper - KELEX 100 system has perhaps been more successful and authors have apparently not needed to resort to empirical analysis. One of the original works in this field, by Spink and Okuhara (347), produced the data on which later models were based, but its explanation of the reaction chemistry relies upon the existence of an almost hypothetical organic species of which there is no real proof. Hoh and Bautista's model (346) is based on the limited data of Spink and Okuhara and again suffers from the need for new parameters for each constant acid isotherm. They show that the equilibrium constant is dependent on ionic strength and sulphate concentration and derive a model which is similar in form to the later work of Bauer and Chapman (348). These workers produced a considerable amount of data from a statistically designed laboratory campaign and their model, although empirically derived, utilizes a thermodynamic reaction equilibrium constant and activity coefficients. It is fairly compact and with a standard error of 9% has been used in design work.

As has been shown, the modelling of copper liquid-liquid extraction systems has to some measure been successfully achieved. Although this has only been on a laboratory scale the reagents used have been found to scale-up adequately to large scale plants (341, 345) and so are of use in design; it should, however, be remembered that impurities in plant liquors may have a significant effect on performance and that some testwork will be required to determine this.

2.6.4 Summary and Conclusions

Liquid-liquid extraction techniques are playing an increasingly important role in various hydrometallurgical processes. There are numerous chemical reagents and contacting equipment types available for use in such processes and the task of designing a suitable system can be complex, taking into consideration a number of important variables and choosing an appropriate reagent-equipment combination.

Whilst many systems would require extensive laboratory and pilot scale testwork before design work could begin with any confidence, processes using known reagent and equipment combinations need considerably less parameter determination. Other than confirmatory tests to study the effects of solution composition such as impurities, the viability of processes such as copper and uranium extraction can initially be evaluated from existing information and experience, unless

it is intended to use new reagent-equipment combinations.

However generally applicable, theoretical relationships describing even the most commonly used and thoroughly studied systems are unavailable. Indeed the underlying reactions are not fully understood in many systems and so the most useful relationships are empirically based. This is especially true when studying extraction kinetics, and therefore stage efficiency, which are highly dependent on the mixer characteristics; for a widely useful model it is most likely that stage efficiency would need to be determined separately, probably based on experience. Similarly the phase disengagement rate is affected by many variables and very few reliable relationships have been published; this is another system parameter whose value is likely to be defined by an experienced design engineer. However extraction equilibria, the third important system variable, have been modelled successfully, although the most suitable models are empirical in form as the more theoretical models are based on chemical reactions which are still not yet fully understood and have excessively complex mathematical forms. This empiricism has not been found to be restrictive though, and similar relationships have been used successfully in process design studies; two equations describing the extraction of copper were derived from published data. The major drawback of the published models and data is that almost all studies have concentrated on the extraction process, ignoring the importance of the stripping stages; until such studies have been completed it will be necessary to significantly extrapolate the models for extraction equilibria. Similarly the quantitative effect of solution components other than copper and acid are largely unknown and ignored.

Once equilibria relationships and stage efficiencies have been established for a particular system, a mass balance around the process can be obtained relatively easily, even for the commonly used counter-current circuits, using computational techniques. This is more accurate than traditional graphical design methods, as well as being faster and more convenient.

Table 2.2 : Operating characteristics of solvent extraction plants (257)

	Ranchers	Bagdad	Nchanga	Cities Service	Anamax	Ray Mines
Solvent	LIX64N	LIX64N	NIX64N	LIX64N	LIX64N	P5100
Carrier	Kerosene	Napoleum	Escaid	Napoleum	Kerosene	Escaid
Organic flow (m^3/min)	8.7	12	52	11.4	38.3	34.0
Aqueous " (m^3/min)	7.6	12	48	11.4	23.5	22.7
[Cu] (kg/m^3)	2.0	1.1	4.5	0.9	3.0	3.0
[H_2SO_4] (Kg/m^3)	3.0	1.9	1	1.0	1.9	10.0
Raffinate [Cu]	0.15	0.1	0.3	0.1	0.13	-
Extraction stages	3	4	3	3	4	2
Stripping stages	2	3	2	2	2	2
Mixer residence time (min)	3	3	3	3	3	2
Settler residence time (min)	6	3	14	13	3	2.5
Electrolyte flow	3.0	0.5	10	0.64	2.8	11.4
[Cu] in	35	25	33	30	23	40
[Cu] out	42	50	60	50	46	-
[H_2SO_4] in	155	185	175	160	-	170
[H_2SO_4] out	145	150	145	135	140	-

2.7 Copper Electrowinning

One of the most common methods for the recovery of copper from leach solutions, this process entails the application of an electrical potential between an inert anode (often a lead alloy) and a copper cathode, both immersed in the copper bearing solution, thus depositing more metal onto the cathode. In 1974, over 550,000 tonnes of cathode copper were produced in this manner (257).

Assuming that sulphuric acid was used as the lixiviant in the leaching stage, the overall electrowinning reaction is:-

$$Cu^{2+} + SO_4^{2-} + H_2O \rightarrow Cu^{o} + \tfrac{1}{2}O_2 + 2H^{+} + SO_4^{2-}$$

As this equation shows, the products of the processes are copper metal at the cathode, oxygen at the anode and a regeneration of sulphuric acid for recycling; it is this last property of the reaction which makes electrowinning so useful, as only minor acid additions are required, to compensate for any losses through spillage and evaporation, and a closed loop circuit can be used reducing the release of harmful agents to the external environment.

2.7.1 Copper Electrowinning Plant Design[1]

Present day copper electrowinning is based upon electrorefining technology, the origins of which reach back well into the 19th century. Experience and understanding of the processes have obviously built up over the last 100 years, although the basic techniques have changed little; plant design is still very similar to the original tankhouse concept of a series of rectangular tanks built end to end with common walls.

The underlying principles of electrowinning plant design and operation are Faraday's laws. These can be stated as: "the passage of 96487 coulombs through an electrochemical reactor produces in total, one gram equivalent of products at an electrode". In the case of a single reaction occurring at an electrode:-

$$n = \frac{I\ t}{z\ F}$$

---

(1) Due to the lack of available information on electrowinning plant design and the repetition of standard formulae in the following section references are not quoted exhaustively; the majority of the discussion was developed from textbooks (257, 364-367) and technical papers (368-378) as well as those quoted in the text.

where n = Wt of metal deposited

I = Deposition current

t = time

z = ionic charge of metal

F = Faradays Number

= 96,487

If it is accepted that almost all copper in solution is eventually recovered in the tankhouse, the design problem is then defined by the deposition of a particular tonnage of copper, this being determined by the amount of copper produced in leaching; therefore the total plant current is constrained by the equation describing Faraday's laws.

The designer's task is thus to plan a system to produce the required amount of copper, using this current, with minimum investment and operating costs; other than tonnage and current, the major system variables affecting plant design and economics are applied voltage, current efficiency and current density.

The total applied voltage across an electrowinning cell may be divided into the following components:-

1) Electrode potential (E)

2) Electrode over-potentials (η)

3) Potential drop due to ohmic effects

For the deposition of copper from a sulphuric acid solution, the theoretical cell potential (E^o) is 0.89 V, and from this the theoretical decomposition potential (E) is given by:-

$$E = E^o - \frac{RT}{zF} \ln \frac{(a_{H^+})^2}{a_{Cu^{2+}}}$$

where T = temperature

a_{H^+}, $a_{Cu^{2+}}$ = activities of ions in solution

If it is assumed that the hydrogen activity in the strongly acidic conditions prevailing is unity and that cuprous activity is 0.1 (i.e. approximately 0.5 M solution), the theoretical decomposition potential is given by:-

$$E = 0.89 - 2.303 \frac{RT}{nF}$$

The electrode overpotential component of cell potential (η) largely comprises of the anode overpotential as the overpotential for the cathode deposition of copper is relatively small (approximately 100 m V). In electrowinning from sulphate solutions, the anode reaction

is the evolution of oxygen and is well known for its irreversibility, especially when using lead anodes with lead dioxide present on the surface. The anodic overpotential can normally be calculated using a Tafel "law" relationship, these being empirical equations describing the dependence of η on current density; these are commonly used and available for most important reactions, and are of the form:-

$$\eta = a + b \log i$$

where i = current density

a, b = empirical constants

The third cell potential component, due to ohmic effects, consists largely of the potential drop across the electrolyte; this can be found from:-

$$V_{RE} = I \ell d$$

where V_{RE} = potential drop due to electrolyte resistance

I = cell current

ℓ = electrolyte specific resistivity

d = distance between electrodes

This potential drop can be quite considerable relative to cell voltage (approx. 0.5 V) and is also dependent on electrolyte temperature and composition; empirical, but generally applicable, relationships for determining conductivity have been given recently by Price and Davenport (349), allowing determination of this potential drop under a wide range of conditions. The effect of the resistance of the electrode and other cell hardware is considerably more difficult to estimate, due to the variation of current density over the electrode, variation of electrode thickness and the difficulty of measuring the effect of busbars, connections etc.; however a value of about 100 m V is a reasonable approximation when considering a modern, well managed tankhouse (350).

Therefore the total cell potential required can be found from:-

$$V_T = E + \eta_A + |\eta_c| + V_{RE} + V_{RH}$$

or $$V_T = (0.89 - 2.303 \frac{RT}{nF}) + (a + b \log i) + (0.1) + (I \ell d) + (0.1)$$

As can be seen, the cell potential is highly dependent on electrolyte temperature and, more significantly, current density; generally, with an increase in current density, voltage, and therefore power, requirements also increase significantly.

The second major system variable, current efficiency, is defined as the fraction of the actual cell current used for depositing copper. The major current losses arise from short-circuiting, stray currents and the reduction of ferric ions at the cathode. The effects of short and stray currents are difficult to predict, but are often of the order of 3-4%. Losses due to ferric reduction are also difficult to estimate, but may be limited by minimizing iron content and operating at high current densities. However current inefficiencies in copper electrowinning usually only vary within a fairly narrow range and can be reasonable reliably predicted for given operating conditions, based on experience.

Although perhaps not critically important with respect to production performance and material balance considerations, current density is often crucial to the economics of a plant. With the level of production depending on total current according to Faraday's law, cell size will depend on current density and so to reduce capital costs this would need to be maximized. As can be seen from the equations describing cell potential though, if the current density is increased, voltage requirements will also be higher, resulting in a significantly larger power consumption.

For the majority of electrolytic plants, the design task is to optimize profitably by balancing capital and operating costs; mathematical means of calculating this optimization have been published (351-353) and a study by Balberyszski and Anderson in 1972 (354) predicted an optimum current density of 671 A/M^2 for copper electrowinning. It would appear that the optimum is now in the range of 600 - 1000 A/M^2 depending on local conditions and prices. However these densities are unattainable with present technology mainly due to problems of deposit quality and so plants tend to be operated at the maximum that quality and/or quantity restrictions allow.

Electrode kinetics in standard parallel plate electrowinning cells are mass transfer processes and are controlled by diffusion through a boundary layer between the metal surface and the bulk solution. If the concentration of active ions in this layer is inadequate for the rate of electron transfer (i.e. current density) the diffusion rate will fall; this in turn results in rough or powdery deposits at the cathode and increases the layer of polarization at the anode. To guard

against this and to ensure adequate current density it is essential to maintain the concentration of metal ions. This can be achieved by increasing the bulk copper concentration in solution, increasing ionic mobility through raising the temperature or introducing depolarization ions such as chloride.

Alternatively, consideration of Fick's Law and the Nernst model shows that, rather than increasing the mass transfer driving force, the rate of diffusion can be improved by reducing the boundary layer thickness. Recent innovations in this area have been fluidized bed electrodes, agitation at the electrode face and periodic current reversal. Although studied by a number of workers (355, 356) and certainly of interest for the treatment of very dilute solutions, fluidized bed electrodes have remained more of a concept than reality with respect to leach liquor processing and have not yet progressed from pilot testing. This is also true for periodic current reversal techniques, although such a system has been studied on an industrial scale in Zaire (357). However it would appear that operating problems have not yet been resolved and that the economic implications of the considerably higher electrical power consumption are particularly dubious, especially in the present energy climate. The third innovational area, that of increasing the mass transfer rate by agitation of some form, has however brought some changes in cell design and operation, although at the moment it would appear that mechanical and ultrasonic methods are not cost efficient. However the improvement of flow characteristics and/or air sparging techniques are worthy of further consideration. Large enough fluid velocities to create any significant thinning of the boundary layer are virtually impossible in conventional cells and so development of new cells has been undertaken at various establishments (358-360). The only cell to have been tested and used in a production environment, the CCS directed circulation system, is thought (361) to incur such high pumping costs that any benefits from its use in copper electrowinning are marginal. However recent work at Kennecott by Harvey et al (362, 363) has shown that air sparging could have a significant future in electrowinning tankhouses, achieving densities of up to 2000 A/M^2 with conventional cells fitted with very basic auxiliary equipment and only low extra power consumption and producing high grade, cathode quality deposits even with low solution concentrations.

It would therefore appear that for the near future there is unlikely to be any major changes in cell geometry from the established designs, but that minor alterations may be made, some of which could have significant effects on operating parameters.

Even without allowing for such changes in operating parameters, the prediction of the behaviour of just a simple parallel plate reactor from theoretical considerations is impracticable. Most of the relationships describing the deposition of copper have been determined with reasonable reliability; indeed the use of these for cell design has been comprehensively covered in a recent monograph by Pickett (364). However some of the important inter-relationships, for instance between cathode quality, solution and deposit impurities and current density, are still largely unquantified and need to be measured for particular operating conditions using pilot scale equipment. Thus to analyse an electrowinning plant on a cell by cell basis, incorporating micro-scale characteristics, and to keep abreast of developments in cell design, a designer would need to use quite detailed but empirically derived relationships which are only obtainable through fairly extensive testwork. Although this work would probably be essential for detailed planning and design, it is probably that much of it would be superfluous to the requirements of an initial process evaluation.

In comparison with published data from operating plants, it can be shown that for modern plants, treatment of the operation simply as a black box, only considering input and output flows, and assuming reliable prediction of operating parameters such as current density and current efficiency, can provide fairly accurate estimates of fluid flows and concentrations, electrode requirements and power consumption; examples are shown in Table 2.3. The calculation procedure only requires the use of Faraday's laws, stoichiometric relationships, and simply derived equations. Whilst considerable plant input from the design engineer is required and a great reliance is placed on his experience, this type of model is simple and robust, and does not rely on experimentally determined parameters, as well as giving flexibility to allow for cell design improvements. From the six examples shown in Table 2.3, it can be shown that the maximum relative error in the calculation of power consumption is only 11.7% and the maximum area for electrode area is only 4.5%.

Table 2.3 Electrowinning Plant Variables and Relationships

Mine	Reference	Annual Cu Prodn. rate (tonnes $year^{-1}$)	Current Efficiency	Plant Current[1] (A)	Calc. Electrode[2] Area (m^2)	Actual Electrode Area (m^2)	Current Density (A/M_2)	Calc. Potential[3] (V)	Actual Potential[4] (V)	Calc. Power[4] (kWh $tonne^{-1}$)	Actual Power
Cyprus Bagdad Arizona	352	6500	89%	733,587	3762	?	195	2.07	2.0	1962	2200
Bluebird Arizona	1	6800	90%	758,917	3035	3110	250	2.19	2.0	2054	1900
Cities Service Arizona	1	5500	90%	613,830	2360.9	2367.6	260	2.33	1.9	2187.9 (DC)	1870 (AC)
Chambishi	1	18000	85%	2,127,070	7734.8	8086.4	275	2.5	2.5	2481	2700
Nchanga I	1	80000	88%	9,131,363	28,535	?	320	2.6	2.0	2492	2200
Nchanga II	1	96000	85%	11,344,376	56,721	57,760	200	2.29	2.25	2273	2400

Calc. Free [5] H_2SO_4 output ($kg\ m^{-3}$)	Actual Free H_2SO_4 output
191.7	185
156.8	151
139.0	160
35.0	30.0
188.0	185.0
55.23	55.0

Notes

1 Theoretical Current to deposit Copper (I_{Cu}) $= \dfrac{\dfrac{\text{Annual tonnes}}{63.54} \times 2\ 9650}{350 \times 24 \times 60 \times 60}$ (from Faraday's laws)

Plant Current $= \dfrac{I_{Cu} \times 100}{\text{Current Efficiency}}$

2 Electrode Area $= \dfrac{\text{Plant Current}}{\text{Current Density}}$

3 Potential $= E + \eta_A + |\eta_c| + V_{RE} + V_{RH}$

4 Power $= \dfrac{\text{Potential} \times \text{Current} \times \text{Time}}{\text{Production}}$

5 $[\text{moles Cu}]_{in} + [\text{moles } H_2SO_4]_{in} = [\text{moles Cu}]_{out} + [\text{moles } H_2SO_4]_{out}$

2.7.2 Summary and Conclusions

Over the last century there has been little change in the design and operation of copper electrowinning tankhouses, although there have recently been experiments to increase current densities by improving cell design. Whilst there has also been some research into improving the understanding of the underlying processes involved into the electrode deposition of copper, this has not resulted in major advances in the methods used to design tankhouses.

It would appear from the limited available literature on electrowinning plant design that the detailed design of these plants involved extensive pilot-scale testwork on the type of solution likely to be encountered in operation. However it is unlikely that this type of testwork is carried out for initial process evaluations and it would appear from analysis of six operating plants, that it is reasonable to consider the tankhouse as a integral unit operation and to predict operating parameters from Faraday's Laws, semi-empirical relationships to determine cell voltage and stoichiometric relationships. Whilst it is necessary for the user of such a model to select feasible values for current density and current efficiency, it reasonable values are used then the relative accuracies of less than 12% can be achieved. The equations used are shown in Table 2.3 and the detailed form of the relationships used to determine cell voltage is discussed in the preceeding text. It it is assumed that the main task in designing an electrowinning tankhouse is to plan a system that will produce a required amount of copper annually, efficiently and economically, and that the major constraints on designing such a system are the practical considerations of what operating voltage, current density and current efficiency might be achieved, then it is reasonable to initially evaluate such a plant using such simple but robust relationships.

2.8 Physical Separation Processes

2.8.1 Gravity Separation

The difference between the specific gravities of two minerals provides the basis for one of the most commonly used methods of mineral separation. Gravity separators are used in many roles, from pre-concentration to concentrate up-grading, and come in a multitude of machine types, the more important ones including jigs, heavy media separators, tables, spirals and cones; Mills (379) has shown that in the United States the value of products from gravity separators was treble that from flotation processes.

However examination of the literature of the past 15 years shows the remarkable neglect of gravity separation, reflecting the sparse research that has been carried out into the subject (379); it should also be noted that those papers that have appeared have tended to concentrate on description of a process application or a new machine, notably the Reichert Cone, with few discussions on the principles of the processes.

The approach to the design of a gravity separation circuit is fairly straightforward (379, 380), with a heavy reliance being placed on the interpretation of heavy liquid analysis data, which can easily be carried out in a laboratory and which requires only a small amount of sample. Whilst the choice and design of separation machines can usually be made on the basis of the analysis results and a knowledge of the particle size distribution and throughput, with perhaps some confirmatory testwork on small laboratory separators, the calculation of a detailed circuit mass balance would seem to be dependent on the estimation of a machine's separation characteristics for a particular feed material, information which is often supplied by the equipment manufacturer.

As was implied earlier the development of phenomenological models of gravity separation processes is almost completely lacking, although there has been extensive use of empirical partition curve models in the coal preparation industry. At the heart of all of these models are generalized distribution curves, based on techniques derived by Gottfried and Jacobson (381) and developed for a number of commonly used coal washing devices (382); these curves define a separation by which a given proportion of the feed to the unit either reports to a clean coal or

refuse stream as determined by its mean specific gravity and size. Several programs (129, 383-385) have been developed utilizing these curves and allowing a wide variety of coal washing circuits to be simulated.

2.8.2 Other Physical Separation Processes

Even in comparison with gravity separation, the literature on the use, principles and design of devices such as magnetic and electrostatic separators and sorters is sparse indeed. The only assumption that can be made is that testwork is carried out on laboratory scale machines similar in design to their industrial counterparts and that mass balances are calculated on the basis of separation efficiencies or partition curves describing recovery as a function of a physical property such as magnetic mineral content or radiometric content. The author is unaware of any published model which can be used to simulate such processes.

2.8.3 Partition Curve Modelling

As has been seen some unit operations, such as gravity, magnetic and electrostatic separators, have not been entirely successfully modelled, especially with regards to design purposes. Although the underlying physical phenomena that govern these processes are understood, viable mathematical models have been difficult to develop.

Separation effects for such machines are often described by separation efficiency curves, also known as partition or Tromp curves. These curves show quantitatively the performance of individual particles in separation operations so that each particle, once characterized, can be accounted for and their sum total will give the performance of the unit. The use of partition curves in process design is fairly commonplace, the usual practice being to choose a feasible curve for a type and application of equipment and to obtain a mass balance using this.

Normally the characteristic partition curve of a separation device will vary with its feedrate and so it is necessary to choose a curve suitable for the expected operating conditions, obtain a circuit mass balance and compare the expected with the calculated conditions. If these are significantly different, new curves will have to be used and the mass balance recalculated; the procedure will thus be repeated iteratively until the selected partition curves are appropriate to the

mass balance. The advantages of computerizing the mass balance calculations are obvious and if the selection of partition curves could be similarly automated, design procedures would be greatly enhanced. However, whilst the mathematical forms of some unit operations' characteristic curves have been successfully modelled, their relationship to feed and operating conditions is more uncertain. Despite this, the use of partition curve models makes process simulation possible even when phenomenological models of a unit operation are neither available nor appropriate; the iterative nature of the design procedure also makes computer simulation desirable, if not essential.

By convention it is assumed that the selection of process material of a certain type is first-order with respect to the quantity of material of that type presented in the feed to a process and that the proportionality constant for the higher-order exit stream (e.g. coarser or heavier) is defined as the process selectivity value, s(x):-

$$s(x) = \frac{\text{Weight of material of property x in 1st exit stream}}{\text{Weight of material of property x in feed stream}}$$

The proportionality constant for the 2nd exit stream (fines or light) is 1 - s(x). When the selectivity values are plotted against the relevant property, it is assumed that the points lie along a continuous functional form; this is known as the separation efficiency curve, selectivity function, position curve or Tromp curve. They are used to characterize the effects of one physical property on an otherwise homogeneous feed passing through a separation machine; thus it can be used to describe the size separation achieved by a classifier on an homogeneous feed. Alternatively, a material can be described by discrete distributions of one property and the effects of a second property described by a partition curve for each discrete fraction; in this way the effects of size can be allowed for in a gravity or magnetic separator.

Ideally a separator would achieve perfect separation about a certain property as in Fig. 2.19; an actual partition curve is shown in Fig. 2.20, the deviation from the ideal representing the imperfection of separation. The differences denoted by a and b in Fig. 2.20 represent the arrival of particles in exit streams independently of the separation property; these bypass fractions can be excluded by the use of a corrected partition curve, as shown in Fig. 2.21, and a corrected selectivity value, $s_c(x)$.

Corrected partition curves can be determined, usually empirically, for many separation devices and it is these that a plant designer would use to obtain a circuit mass balance. A general model of a solids separator should therefore be based around typical partition curve types. These could take the form of assuming selectivity values for specified discrete fractions as in Gottfried's coal preparation plant simulator (129); these values can then be interpolated for the actual feed distribution to describe separation. A more elegant solution is to use continuous functional forms to describe the partition curve. This technique has been used by Loveday (386) for gravity separators, Plitt (138), Harris (387) and Luckie and Austin (388) for classifiers and Rose (128) and Brereton (389) for screens. These are generally two parameter functions dependent on x_{50} (see Fig. 2.20) and a measure of the separation efficiency. They are written in terms of s(x) and I where:-

$$s(x) = (x - \lambda)/(x_{50} - \lambda)$$ where $\lambda = 1$ for gravity separators and $\lambda = 0$ for others

and $$I = \frac{x_{75} - x_{25}}{2.x_{50}}$$ x_{25}, x_{50} and x_{75} = value of the property at 25%, 50%, 75% recoveries respectively.

The theoretical x_{50} for a material can be determined from laboratory testwork; I is an empirical constant describing the imperfection of a machine and can be determined from testwork on that machine or, less accurately, from equipment manufacturers and/or experience.

This latter approach has been used by Ford (16) in his proposed ore dressing plant simulator; he has implemented four 'typical' curves proposed by the authors listed above to enable a range of equipment to be simulated. Whilst this would appear to be the optimum development route, no results have been published on the general application of such an approach and a wider range of functions may be desirable, especially if this were to be compatible to those used in industry.

2.8.4 Summary and Conclusions

It has proved difficult to usefully model phenomenologically many physical separation processes. In plant design these operations are generally analysed using partition curves which describe their recovery efficiency for different types of material. Although the selection of an appropriate partition curve for the particular operation conditions

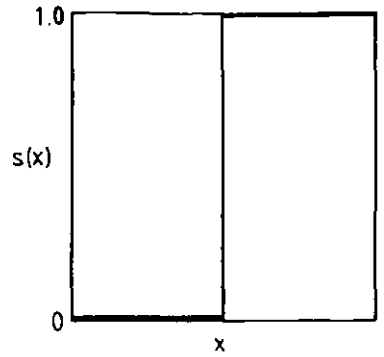


Figure 2.19 Curve describing perfect separation

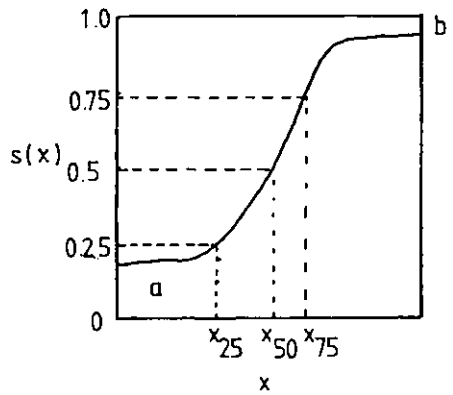


Figure 2.20 A typical partition curve

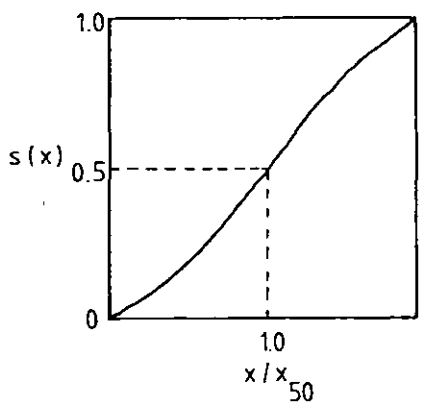


Figure 2.21 Corrected partition curve

is often an iterative process, the availability in a simulator of a unit operation block which can describe and use partition curves to model a process would be of great value. One of the most efficient methods of implementing such a block would be to include a number of optional functional forms of general-purpose partition curves, these being based on those found to be useful in describing a variety of unit operations; to improve accuracy it should be possible to define a separate curve for each particle size range in the feed material. An alternative method is to model particular unit operations using a defined set of curves for that operation; an example of this type of unit operation model is the simulation of different coal washing machines by Gottfried (129).

CHAPTER 3

Process Costing and Economic Evaluation

The economic evaluation of a process involves the estimation of capital and operating costs, the prediction of potential revenues and an analysis of the implications of these factors in the project's profitability. Inherent in this analysis is also the identification of the critical technical and economic parameters of the process, and the presentation of the evaluation results such that the problems of comparison and choice between options are minimized.

A consistent approach is important in using evaluation techniques so that realistic comparisons between options can be made and to ensure that errors of omission, especially through the use of shortcut procedures, are significantly reduced. Although this can be achieved through the rigorous use of flowchart cost model (390), the use of computer based techniques is perhaps a faster and more reliable method; a further advantage of computerizing economic evaluation techniques is the extra efficiency provided for the rapid storage and retrieval of cost and economic data. Utilizing such techniques, there is now a considerable number of programs for use in capital cost estimation (29, 391) and financial analysis (25, 26, 392) packages for combining these functions in an overall evaluation (29, 393) and many subsystems of larger simulation systems performing a similar role (391).

It is worthy of note that these packages tend to be readily accepted in industry, largely because they are usually based around established manual methods and so are not introducing radically new concepts or methods, but are merely improving speed and reliability. A recent study of computer based cost estimation methods (29) claims that although it is virtually impossible to find any convincing evidence of their cost effectiveness, all of the industrial users contacted believed their programs to be worthwhile.

3.1 Capital Cost Estimation

Perhaps the most important step in any project analysis is the estimation of the required capital investment; this investment will include the fixed capital cost of equipment and buildings, land purchase and provision for working capital.

The fixed capital cost is commonly regarded as consisting of:-

1) the purchase and installation of equipment and all ancillary

services for that equipment

2) the provision of buildings and administrative facilities

3) contractor's fees, expenses and contingencies.

Working capital is generally understood to be made up of the funds, in addition to fixed capital and land investment, which a company must inject into a project to enable it to start operating and to meet subsequent obligations as they come due (394).

Fixed Capital Cost

Just as there are various levels and classifications of cost estimates (see Chapter 1) dependent on their accuracy, so there are also various estimation techniques which may be similarly classified; three classes commonly used are the shortcut, factorial and detailed methods.

"Shortcut" methods

"Shortcut" methods are usually used in the preliminary stages of a project when only the type of plant and its capacity are known and an order-of-magnitude estimate is required. In general they are of most use when investigating new process routes and rely heavily on a large source of data (29). Despite the rudimentary nature and known inaccuracies of these methods, there has been a considerable amount of study of this field recently; this includes the work of Zevnik and Buchanan (395), Gregory and Bridgwater (396), Stallworthy (397) and Wilson (398), the bulk of this study having been reviewed and used on a comparative basis by Allen and Page (399). Five methods were used to estimate the costs for six different types of plants at various capacities whose real values were already known; the average errors for each method were found to vary between 14% and 63%. However use of these methods by different workers resulted in vastly different average errors (29) highlighting the empirical nature of these methods.

One type of shortcut estimating method is based on the number of "functional" units in the process flowsheet, a measure of the complexity of the plant being obtained from and related to the maximum operating temperature and pressure and the likely materials of construction.

Thus:-

$$\text{Plant cost} = \text{Constant} \cdot \left(\begin{matrix}\text{Number of}\\ \text{Functional}\\ \text{units}\end{matrix}\right)^{\text{Index}} \cdot \left(\text{Capacity}\right)^{\text{Index}} \cdot \left(\begin{matrix}\text{Pressure}\\ \text{Function}\end{matrix}\right)^{\text{Index}}$$

$$\left(\begin{matrix}\text{Temperature}\\ \text{Function}\end{matrix}\right)^{\text{Index}} \cdot \left(\begin{matrix}\text{Materials}\\ \text{Factor}\end{matrix}\right)$$

The constant and indices are found by regression of available data (395, 396).

A more basic but also more commonly known estimation method is known as the "six-tenths rule". When historical data is available on a similar type of plant, it has been found that the cost of a new plant of differing size is proportional to the ratio of the relative capacity raised to a power.

$$\text{Cost new plant} = \text{Cost old plant} \left(\frac{\text{Capacity new plant}}{\text{Capacity old plant}}\right)^{X}$$

X has been found typically to vary between 0.5 and 0.8 for overall plant costs, though this technique can also be used for determining the cost of sections of a plant, in which case the value of X can be outside this range. For many chemical processes, regression equations have been developed and published relating cost directly to capacity and/or other technical factors.

Whilst these methods are useful during the initial stages of project development and in screening potential processes, it should be remembered that they produce global cost estimates and that great care is necessary in their use. Almost all process plants, whilst similar in nature, differ in some way; this can be in the ease of development, different ancillary provision, different project managements and different construction standards. The methods also make no provision for cost escalation over time nor for the effects of plant location.

In general, this type of cost estimation method is used in situations where only basic technical information is known about the process, and the forms of the relationships on which they are based reflects this lack of knowledge; this is also reflected in the accuracies which are claimed for this type of estimation, usually between $\pm$ 30% and $\pm$ 50%. They are therefore unsuitable for the type of process analysis proposed where more reliable estimates are needed and where fairly detailed

information is available on the plant flowsheet and its constituent equipment.

Detailed methods

Detailed capital cost estimation techniques are usually used when a particular process route has been chosen and equipment specifications are being drawn up. Rather than estimating the probable proportion of equipment types and building materials that will be needed, fully defined plans of what is necessary are drawn up and price quotations obtained from equipment manufacturers. Highly reliable and detailed cost estimates are therefore drawn up and are likely to provide the basis for cost control in the construction stages of a project. Due to the information and effort required to produce such estimates, they are rarely used before the later stages of a project. The use of such estimates in combination with a process modelling analysis is precluded unless very highly reliable models and sophisticated plant design routines are available and a large effort is expended in obtaining cost information.

Factorial methods

Probably the largest class of plant capital cost estimating techniques is that encompassing the various factorial methods. These are widely used in situations where the basic process flowsheet, including basic equipment sizes, is known and a fairly rapid but approximate estimation is required. The factorial method produces an estimate of the installed cost of the whole plant by multiplying the delivered cost of the major items of process equipment by a number of, or a sum of, factors. The use of factors allows for those costs likely to be incurrent over and above the price of the basic process equipment; these include items such as installation, piping, instrumentation, provision of utilities and process buildings. The many different variations on the basic concept have often been described in the literature (e.g. 29, 30, 32, 394, 400-403); these descriptions also include some useful examples of the typical values of factors.

The complexity and flexibility of these variations increases from the plant cost ratio method, which uses the total equipment cost and a single factor depending on the type of the process, to the module method, which applies a range of factors to each piece of process equipment, before summation and application of further factors to produce the total plant cost estimate. Generally the equipment cost for each

process unit operation is found from correlations of capital cost with one size or capacity specification; outside of unpublished "in-house" correlations, it would appear that there has been no attempt to correlate simultaneously cost with more than one design parameter, thereby assuming that strict geometric and material similitude exists throughout an equipment category. Whilst this may appear excessively rigid, this method of estimating equipment cost is widely practiced and is useful as long as the category definition is not too general and the listing of included equipment is strictly adhered to and observed.

The crudest factor method amounts to:-

$$FCI = Factor \cdot \sum_{1}^{j} \{MPI\}_j$$

where FCI = plant fixed capital investment
MPI = main plant item cost
j = no. of main plant items.

Typical values of the factor are 3.1 for solid process plant and 4.74 for fluid process plant (401) as with all factors used in this type of estimation method, these values are totally empirical and will vary amongst even similar plants, but it should be noted that using only one value to allow for all of the extra plant costs other than equipment purchase is fairly inflexible and is likely to produce erroneous results.

A similar method which again uses only one factor value for all extra costs is the equipment cost ratio method; in this each category of equipment is treated separately with a corresponding factor for each. Thus:-

$$FCI = \sum_{1}^{j} F_j \cdot MPI_j$$

where F_j = factor corresponding to equipment category j.
This method is slightly more reliable and flexible than using one global factor, but still relies on considerable experience of costing an experiment category and requires expertise when allowing for local variations in the design details of a process plant.

Whilst still requiring experience to produce reliable estimates, more detailed factorial estimation techniques allow more flexibility by breaking down those additional costs above equipment purchase into components whose costs are a ratio of major equipment costs; examples of these components are process piping, instrumentation, buildings and

site development and engineering. In this way the cost estimator can make allowances for specific variations in a particular process plant; using the techniques, as the knowledge about a process design increases it should be possible to produce more reliable cost estimates. A number of these techniques are formulated below:-

$$FCI = (1 + \sum_{1}^{i} f_i) \sum_{1}^{j} \{MPI\}_j$$

$$FCI = (1 + \sum_{1}^{k} f_k).\{(1 + \sum_{1}^{i} f_i) \sum_{1}^{j} \{MPI\}_j\}$$

$$FCI = \sum_{1}^{j} (1 + \sum_{1}^{i} f_i) \{MPI\}_j$$

$$FCI = (1 + \sum_{1}^{k} f_k) \sum_{1}^{j} (1 + \sum_{1}^{i} f_i) \{MPI\}_j$$

The flexibility, and potential reliability, of these methods increases sequentially as the sum of factors is applied to the sum of equipment costs, then a different sum is applied to each equipment category in turn; there is also the choice of whether all additional costs are calculated as a proporiton of equipment costs or whether certain costs specific to each equipment category, such as piping and instrumentation, are calculated and the effective total equipment cost for the plant determined before costs related to the overall plant, such as plant utilities and auxiliary buildings, are estimated. Whilst all of these methods are in general use it would seem that the techniuqe most suitable for a modular computer process simulation and analysis system as proposed in this thesis is that which allows flexibility in calculating the total capital costs associated with each unit operation and which then calculates the extra plant capital costs as a proportion of the sum of these.

Once the process simulation is complete and the major equipment items sized, a cost module would be used for each unit operation to determine the purchased equipment cost using an empirical equation relating that cost to a feature of the equipment size or simply to its flowrate capacity. A series of factors would be included in each cost module to allow for installation, piping or conveyors, pumps, electrical and instrumentation equipment etc. The total equipment cost for the plant would then be determined and the total plant fixed capital cost would be estimated using factors for provision of plant services, site development and construction, engineering services and management,

ancillary buildings, etc.

Various claims are made for the accuracy of factorial estimation methods, with values ranging from ± 30% to ± 15%. Whilst a method similar to that outlined above should be nearer the lower end of the accuracy range, it is very difficult to place a figure on such an estimation. The technique relies on equipment sizes which are often determined as a result of the mass balance predicted for the flowsheet; unless a figure can be provided for the reliability of this then it is relatively futile to attempt to predict the accuracy of the cost estimation. Also there has been virtually no data published on where these accuracy figures are derived from nor what they mean.

If a probability distribution of project costs was drawn up it would be a positively skewed (404) rather than a normal distribution as in figure 3.1; the value of the cost estimate and its accuracy limits should in theory be able to largely define the nature of this distribution. It would appear that the relevance of these figures to such a distribution has not been commonly discussed in the literature nor has the value which could be given to the confidence limits for an estimate and its claimed accuracy. In this context it is worth noting that it has been argued (29) that the potential cost overun of a project is higher than the amount by which cost could be under-estimated and therefore it is incorrect to use a figure such as ± x% for the probable estimate accuracy; rather, it is proposed, an accuracy range of + x% and - y% should be used where:-

$$y = 100\ x\ /(100 + x)$$

However there are still no guidelines as to what x and y actually stand for. In a recent paper, O'Hara (404) considers the capital cost estimate to be represented by the mode of the distribution, i.e. the most likely or frequently found value, whilst the median value, on either side of which there is an equal probability of error, is allowed for by provision of a contingency sum in addition to the original capital cost estimate; in this way the recognition that costs higher than the basic cost estimate are more likely than lower costs is taken into consideration.

Working Capital

Whilst Bauman's (394) definition of working capital, as the cash required to get a project started and meet subsequent obligations as

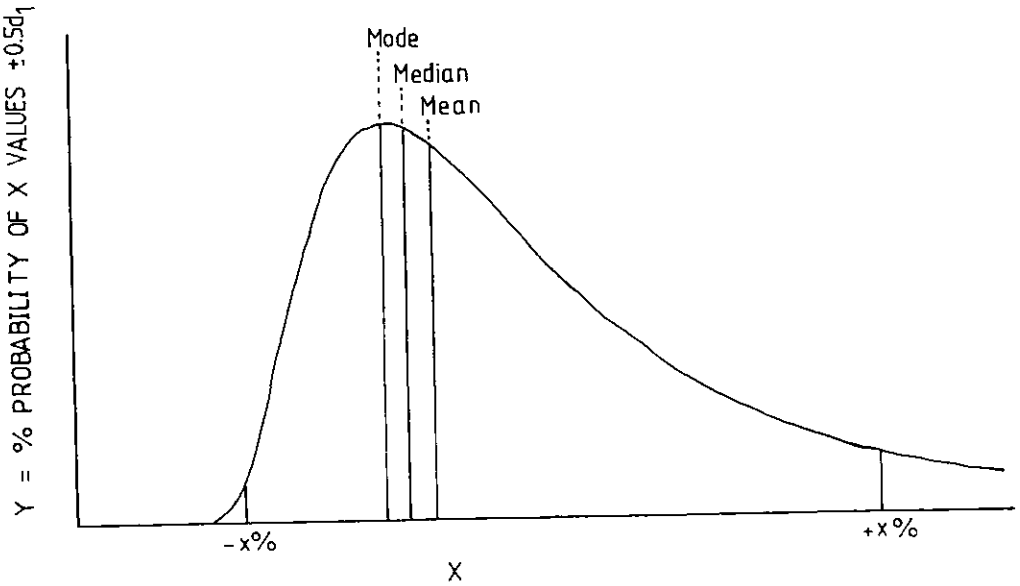


Figure 3.1 Positively skewed probability distribution curve

they come due, may be generally accepted, the extent of its provision and the means of its calculation are wide and varied. As Scott (405) points out, even the engineer's and accountant's definitions are distinctly and misleadingly separate. Although not a particularly tangible concept (406), working capital is typically regarded as consisting of raw materials and stock inventory, accounts receivable and a sum of cash less taxes and accounts payable. Bechtel (407) summarizes the three commonly used methods as:-

(1) the sum of (i) inventory costs = 1 month's raw material
+ 1 month's semi-finished material (raw material + $\frac{1}{2}$ x conversion cost)
+ 1 month's finished product

(ii) accounts receivable (1 month's sales)

(iii) operating cash (1 month)

(2) 30% of annual sales

(3) 15% of fixed capital investment

The latter two are empirical relationships based on experience and range widely, especially between industries. Also, as a percentage of sales, working capital provision will vary depending on the nature of the product and the cost of manufacture (402), whilst in an inflationary environment it is dangerous to relate working capital to a percentage of fixed assets. However method 1 is not totally reliable as there is no allowance for accounts payable, taxes and liabilities, and may also be problematical if there are seasonal fluctuations in costs and/or sales.

It would appear that potentially the most accurate procedure will be to calculate the various components of working capital in a similar fashion to Bechtel's method 1, but in more detail and using typical times for each element as given in Weaver and Bauman (394). Although arduous by manual methods, this would require relatively little effort if automatically computed as part of an estimation program, although for flexibility it would be advisable to allow the estimator the option of inputting his own values.

Capital Cost Data

It can be seen, from the methods described above, that cost estimations are initially based on 'average' costs, calculated from an analysis of historical data, rather than actual quotations from equipment

suppliers. Thus, for well founded estimations to be produced, it is necessary for there to be ready and rapid access to comprehensive and reasonably accurate cost information.

Therefore industrially developed techniques have tended to rely heavily on generalized cost data, usually based on a company's own records of past projects. If this type of information is unavailable or inapplicable then, for the chemical process industry at least, there is a reasonable source in the technical literature. However this is widely spread, not always fully defined and usually related to costs in the U.S.A. With respect to this project, a further drawback is the dearth of such information for the specialized equipment used in the mining and metallurgical industries; although this has been partially alleviated by publications from the CIMM (408, 409) in Canada and the NIM (410) in South Africa, and by less detailed information elsewhere (8, 411, 412), there is still not a particularly wide base of data from which to work and some significant gaps in the equipment covered. There is therefore a need for a comphrehensive data base to be developed; such a data base would be of great service in its own right as well as coupled with the proposed methodology.

Mular, when developing the CIMM cost data (408, 409), used the Marshall and Swift Mining and Milling cost index to update his information. Although this is a widely used index there has been much discussion in recent years on the use of such empirical indices as Marshall and Swift, Chemical Engineering News Record (29, 30, 32, 400, 402, 410, 413-415). Strong arguments have been put forward, especially by Cran (413), for the use of 2 component indices based on the rate of inflation of labour and materials prices; accepting these arguments, the journal "Engineering and Process Economics" have instituted Cran's ideas as a regularly updated feature and have established indices for a large number of industrialized countries. Despite this being a potentially better method however, it would be difficult to apply to data derived using another index and is best used with raw cost data.

A major problem with the use of published cost data is that it is based on prices at a certain location. Therefore there is the need for some technique which allows projects in other areas to be evaluated. Either cost data must be collated for every likely country of interest or there must be a method of using data for one location or country to estimate costs in another. Kharbanda (30) has suggested the use of a

location index derived from the local cost of labour, steel prices and estimated local productivity. The problem is also alleviated by the application of factors to estimate total plant costs, if these factors are split into material and labour components, which can then allow for local differences. It would appear that there is still no satisfactory method for estimating the effects of different locations on costs and so estimates should be based on data derived as locally as possible.

Whilst accurate equipment cost data is important in developing valid plant cost estimates, the dependence of such estimates on correct and accurate use of factors is perhaps even more crucial. Therefore if a cost data base is to develop for a capital cost estimation procedure this dependence should be remembered and accurate examples of cost factors should be collated, although this would be an even more difficult task than the collection of equipment cost data.

3.2 Operating Cost Estimation

The operating costs of a plant are generally obtained by comparison with similar plants and direct analysis of the plant design in question. Thus energy costs may be estimated from the total wattage installed and labour costs from a manning table drawn up by consideration of operating sections, but costs such as administration, supervision and maintenance can only be estimated from previous experience (416, 417). As with capital cost estimation, it is important that all possible costs are taken into consideration and so the standard lists provided in cost estimation texts (32, 394, 400, 403) are helpful.

Operating labour costs are normally derived from consideration of the location and available labour, the type and arrangement of equipment and the instrumentation and control provided. Supervision, clerical, labour, labour charges, payroll, administration and similar labour related costs are then usually estimated as percentages of the operating labour charges. Another option is to total the number of principal processing steps and then using empirical correlations calculate the cost from the man hours per ton and number of steps (403, 418).

Maintenance and repair costs, maintenance overheads and operating supplies may be calculated by developing repair schedules for the various equipment types. This would probably be too detailed for the earlier stages of project development when maintenance related costs are normally estimated as a percentage of capital investment.

As has been stated the electrical power costs are calculated from the installed wattage whilst the cost of other utilities such as water, steam and fuel are calculated directly from requirements. Similarly the cost of consummables such as reagents and comminution media are calculated according to the predicted requirements.

All other operating costs are usually calculated as a percentage of the labour costs, the capital investment or the value of product sales. Indeed, as Bridgwater has shown (418) all of the well known texts can be summarized by the use of a four or five element equation containing percentages of the costs of raw materials, energy, labour, capital investment and, in some cases, sales price. From this study Bridgwater derived a model based on the first four elements and listed typical, maximum and minimum values for the various percentages. In its "typical" form the model is:-

$$1.2\ R + 2.54\ L + 1.2\ E + 0.17\ I = \text{Tot. Op. Cost.}$$

where R = raw material cost
L = labour cost
E = energy cost
I = capital investment

However as a more flexible alternative, it would be preferable to use the more detailed methods and, as with the factors in capital costing, provide a checklist of components and typical values for the respective percentages. Thus the experience of the designer is essential to the methods' accuracy, whilst the rigorous, if not mechanical format, should enhance the accuracy by ensuring that all the components are definitely considered, but only considered once.

3.3 Evaluation of Project Value

Just as there are a number of cost estimation methods in the process industries, so there has also been a proliferation of methods for assessing the financial value of a project. They can be split into two classes, the discounting and the non-discounting methods.

Although of some use in particular circumstances, the latter tend to be the simpler and quicker methods of evaluation, but only achieve this by oversimplifying the often complex problems of investment appraisal (419). They often produce capricious and misleading results

by failing to allow for the incidence of outlays and earnings, and often not considering all aspects of the investment; they are generally regarded as inadequate substitutes for discounting methods whenever there is sufficient time to use such techniques.

Discounting methods differ in that they consider the time value of money and the stage of a project at which cash flows take place. There are two main discounting techniques, net present value and d.c.f. rate of return (or internal rate of return or yield); these are standard techniques and descriptions can be found in the extensive literature (e.g. 32, 400, 419-421). Also in the literature, there are numerous agruments as to the advantages and disadvantages of the two. Whilst there are conflicting opinions about which is superior, there is little disagreement that neither is adequate to be used as an absolute measure for judging profitability. The best solution is detailed study of the annual cash flow whilst using net present value, rates of return and other indices as measures of the project viability.

Whilst the basic concepts and manipulations involved in discounted cash flow techniques are relatively simple, their application is not always so straightforward and, as in cost estimation, can require tedious and repetitive calculations, especially when a long project life is predicted. Also in similar fashion to cost estimation, it is important to consider all possible effects on project cash flow; while some may be difficult to quantify but easy to account for, such as opportunity cost, others can be easy to quantify but difficult and longwinded to account for, e.g. taxation. To develop a valid evaluation on the value of a project therefore requires a consistent and fairly detailed approach involving numerous and repeated calculation procedures, not always understood by economist or engineer, and so the case for integrating these into a computer based methodology is clearcut. As has been said though, whilst the theory is simple the application is not necessarily so and therefore the system used should be as flexible as possible.

The basis of discounted cash flow analysis is to assess the present value of a future cash flow by applying a discount factor to that cash flow; the discount rate used to determine project acceptance needs to allow for the cost of the capital available to the project sponsors and also to ensure a continued cash return, and preferably growth, for the investors. It is in the choice and application of the discount factor

that most problems arise in financial evaluation and so care and understanding is essential. It would appear that there are many interpretations of how this type of analysis should be carried out and so, if only for consistency's sake, evaluations should use those procedures and discount rates in accordance with the normal policy of the organization for whom they are intended; in a recent evaluation of a mine where the government assessed its worth as £40 m a major political story broke when some expert consultants valued the property at £8 m and others at as high as £104 m, the differences occurring largely over confusion about the discount rates (422).

The discount rate used in project evaluation is that rate of return from a project which makes it acceptable to the sponsoring organization. It is generally held that the starting point for finding an appropriate discount rate is to determine the cost to the organization of the capital available for investment; this will vary, not only with the general level of interest rates, but also with the objectives of the company and the risk involved with the projects with which it deals. It is necessary not only to consider the value and timing of project cash flows but also the uncertainty associated with these.

If, as is often the case, the organization's cost of capital is used to determine the discount rate, it is normal to use a weighted average of the cost of capital to the whole organization rather than that specifically used for that project. Whilst this would pay for the capital invested, it does not take into consideration the risk involved in that project, nor does it allow for the continued economic well being and growth of the organization. A risk premium is often used to evaluate the extra return that risky investments must command over less risky investments; however this premium is not easy to determine as it is necessary to predict and numerate the risk in a compound fashion over the life of the project.

A method of considering the generally required rate of return in the financial markets, and which also allows for the risk involved with a type of project, uses a measure known as a beta coefficient. This is a measure of the volatility of individual securities' returns relative to market returns and reflects an industry's characteristics and management policies in determining how returns fluctuate. Beta coefficients can be determined for a particular sector of an industry

by analysis of its rate of return to investors. Having estimated a project's likely beta coefficient, it is possible to represent the expected rate of return for the project as a risk free rate of interest plus a risk premium, where the risk premium is the market risk weighted by an index of the risk of the investment; thus:-

$$\begin{pmatrix}\text{Expected rate}\\ \text{of return}\end{pmatrix} = \begin{pmatrix}\text{Risk free}\\ \text{interest rate}\end{pmatrix} + \begin{pmatrix}\text{Market}\\ \text{return}\end{pmatrix} - \begin{pmatrix}\text{Risk free}\\ \text{interest}\end{pmatrix}\begin{pmatrix}\text{Beta}\\ \text{Coefficient}\end{pmatrix}$$

Typical values for the risk free interest rate can be obtained from the rate on Government bonds, whilst the market return comes from a market index such as that published by the Financial Times. It is important to note that beta coefficients can only be determined from analysis of similar organizations' returns and indeed from the return on similar types of projects and also that this method assumes that analysis of the prices of shares, securities etc. reflects only the risk involved and the necessary rate of return; however other factors such as prospective changes in the structure and control of an organization can have major effects on these prices.

A similar method, though less theoretically based and more difficult to quantify, also splits the expected rate of return into two components, a growth factor and a factor to allow for project risks and sunk costs. The growth factor relates to the risk free interest rate in that it is the minimum rate of return necessary to continue organizational growth and economic prosperity. Similarly the second factor, as in the beta coefficient method, allows for the costs incurred by the organization in investigating and evaluating the project and by those projects which fail to match their anticipated potential.

The discount rate determined by any of these methods is the real rate of return required and should not be confused with the nominal rate of return which is typified by bank lending rates. A real rate does not take into consideration the effect of inflation on project cash flows and viability and so a further step in project evaluation is necessary unless inflation is ignored thereby assuming that the percentage change in annual cash flows is the same as the annual inflation rate. This is a potentially dangerous assumption unless the inflation rate used allows for the different changes in the costs of labour, energy and materials and in the project revenues. A more soundly based procedure is to escalate each of the major types of cost and revenue

separately using appropriate and often different rates, determine the effect of taxation and therefore calculate a nominal cash flow. The real cash flow can then be calculated by deflating the overall nominal cash flow using a more general measure of inflation and the present value of the cash flow then calculated using the real discount rate; alternatively the overall nominal cash flow can be discounted using a rate combining real rate of return and inflation.

Another problem in the application of generalized financial analysis techniques arises in the calculation of annual cash flows and the effect of taxation. The approach to handling taxation in such a generalized technique would have to be flexible to allow for changes in taxation procedures from nation to nation and even between states; for such a technique to be widely applicable a series of tax modules would have to be developed for each of the major mineral producing areas and for other areas of interest.

Summary

The economic evaluation of a process involves the estimation of capital and operating costs, the prediction of potential cash flows and an analysis of the implications of these factors on the project's viability.

The fixed capital cost involved in constructing a process plant is largely dependent on the purchase cost of the major items of process machinery and in a computer based system with the level of detail and knowledge as proposed in this project, the most suitable estimation system is a detailed factorial technique. The necessary working capital would best be calculated from analysis of the inventory, sales and required operating cash. The operating costs would be partially based on the fixed capital costs, on the necessary manning levels and on the predicted energy and material consumption.

Analysis of the project viability would be by discounted cash flow techniques though the actual calculation procedure used would need to be flexible to allow for different organizations' requirements and methods and for different economic conditions. Sensitivity analysis to determine the relative effects of different parameters is an extremely useful advantage of using computer based techniques.

CHAPTER 4

Methodology Integration

Having outlined the proposed methodology, described its potential application and considered in more detail the various components, it is important to discuss and determine how these components can be combined into an integrated system on which the methodology may be based. Figure 4.1 illustrates the stages involved in the evaluation of a process flowsheet, how the methodology handles each stage and the inter-relationship of each component. Whilst an integrated, computer based methodology is proposed, the need for man-machine interaction is anticipated throughout the procedure as there are few areas where the input of an engineer's experience is not considered important, if indeed not vital; therefore any system considered should have the facility for the designer to overide the process model or data and input his own interpretation.

As suggested in Chapter 1 the methodology potentially has many varied applications and it is important to note that the chosen applications can have a significant effect on the program structure. It would be as valid to develop a system to evaluate one, or a very few, commonly used flowsheets, as to attempt to allow a large number of processes to be analysed. The former system would be designed and closely tailored to particular problems and could utilize specially formulated algorithms which were robust and efficient. Whilst this approach necessitates a different program to be written for each application, such programs are useful especially when repeated use is intended. Indeed this was the approach used in the only published similar technique in the minerals industry; Gurun (89) used simple mass and cost models to compare a small number of crushing circuits and choose the optimum configuration. However, despite the difficulties and costs involved in developing a system with the capability, the advantages of general purpose programs capable of handling and combining such evaluations in a variety of configurations have been considered more important by organizations concerned with several different processes (423). Thus for optimum use the methodology needs to be based on a flexible evaluation system.

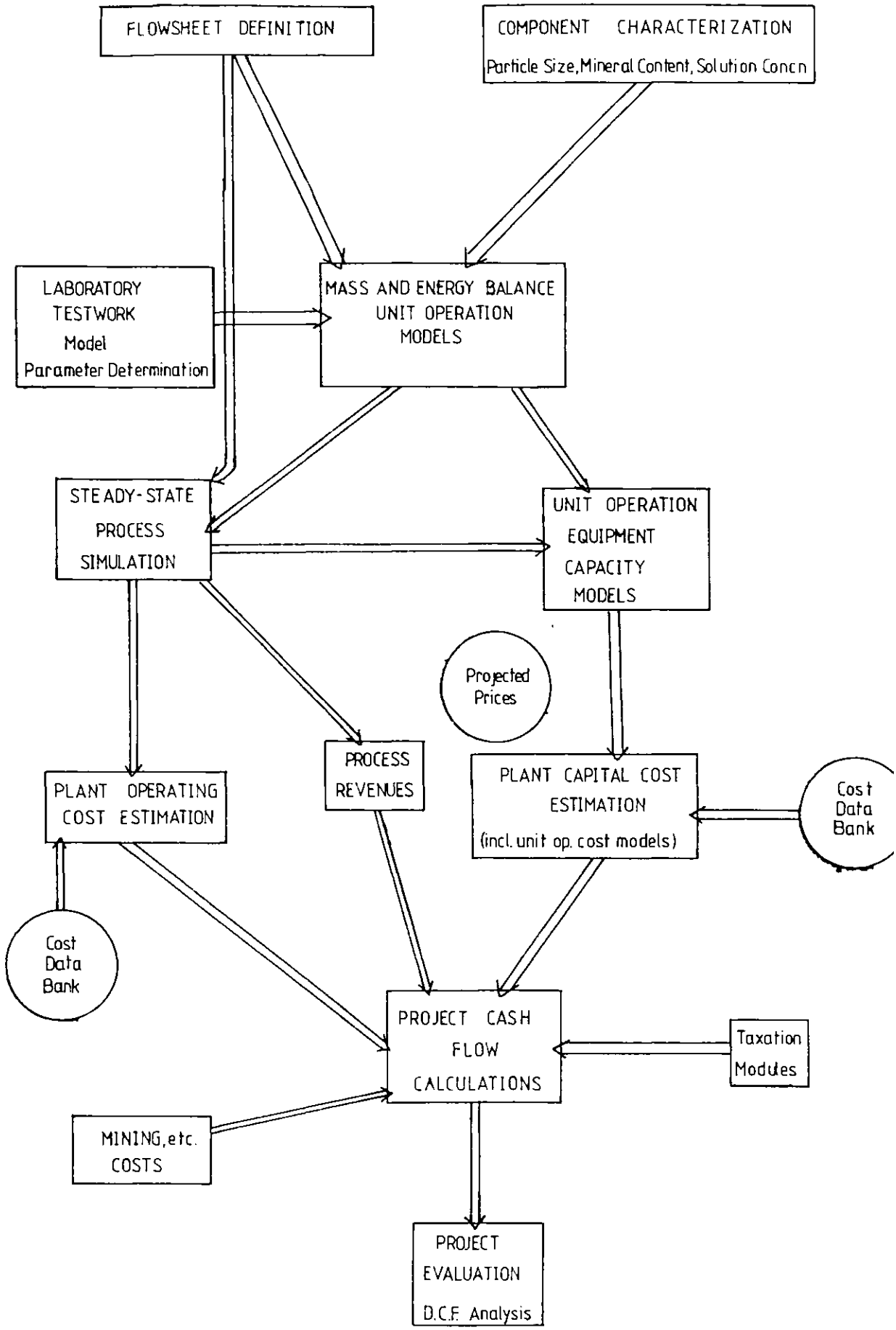


Figure 4.1 Inter-relationship of methodology components

Another important factor in deciding on the system structure is its application and mode of calculation. Whilst the system is intended as the core of a computer-based methodology for process design, there are two distinct flowsheet calculation modes with which to achieve this aim. In the simulation mode all process inputs and flowsheet parameters are defined and a mass and energy balance and process outputs calculated. In the design mode flowsheet parameters would be calculated from constrained process outputs. However it is generally found (27, 424) that in the design mode computational time can be excessive, if indeed a stable solution can be achieved within reasonable limits. If however flowsheet calculations are carried out in the simulation mode, solutions are more readily attained and progress to an optimum design can be achieved in a logical, though iterative, fashion utilizing the designer's experience and an integral learning process which, in turn, can be enhanced using techniques such as sensitivity analysis. Once a simulation model of a process has been developed, it could be used in a controlled series of "experiments" to develop a simple regression model of the process, which could be used more simply to determine the optimum design.

Having decided on these constraints, the system structure can be decided. The most logical way of ensuring system flexibility is to separate the design procedure into simple modules as in figure 4.1, each module containing a separate, basic unit operation model or calculation procedure; thus, if in the possession of an executive module capable of combining these individual modules, it should be possible to develop the most simple or the most involved process evaluation, the only constraint being the available computer resources. Whilst it can be argued (425) that integration of equipment sizing and costing calculations with the mass and energy balance computations is important, this is only likely to be of use in the design mode, and a more logical procedure would follow that used manually: i.e. flowsheet mass and energy balancing → equipment sizing and costing, → operating cost and revenue estimation → cash flow calculation → project evaluation.

A suggested system structure with calculation procedure is shown in figure 4.2. As can be seen it consists of four main parts:-

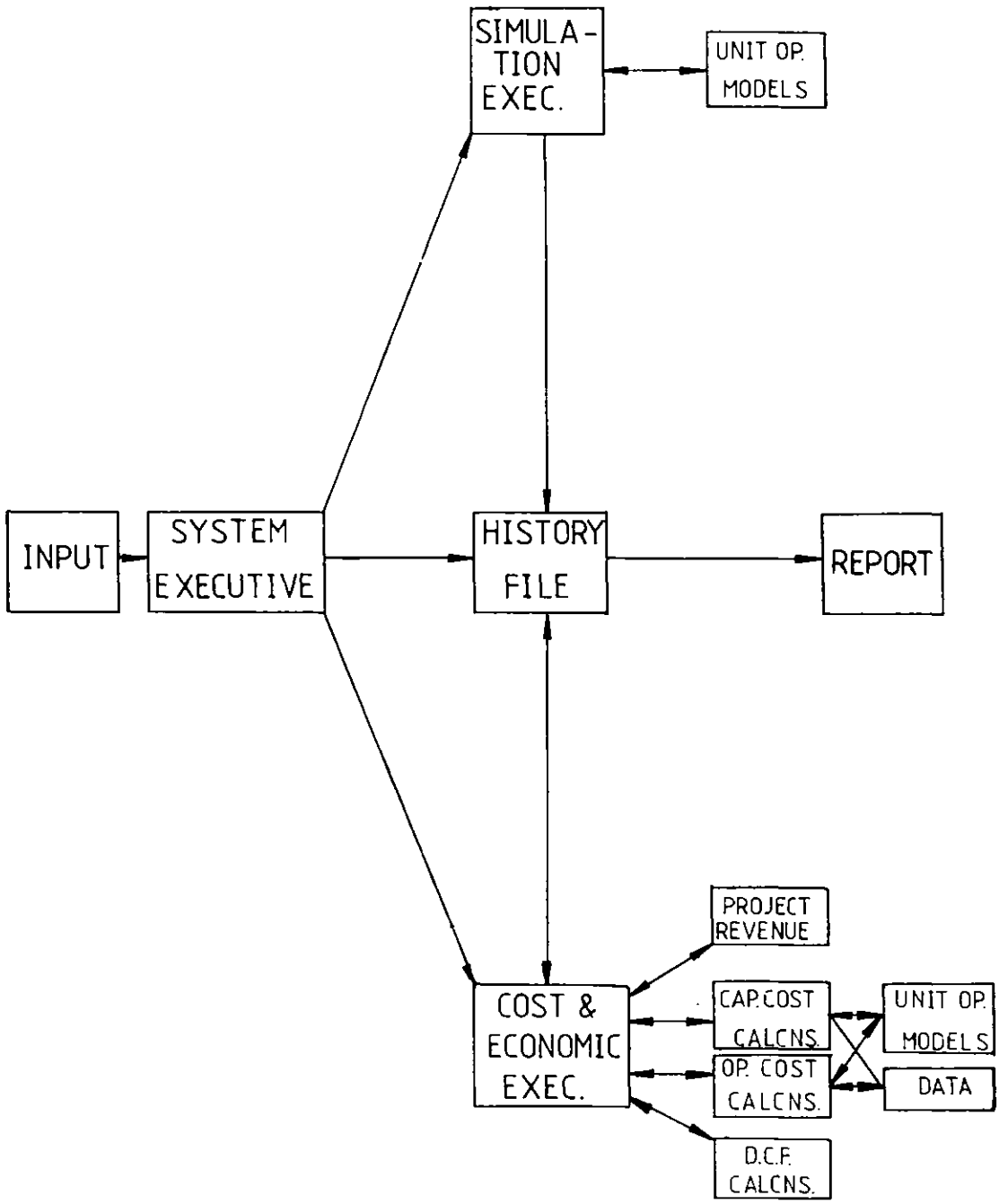


Figure 4.2 Schematic diagram of a computer system for the methodology

1) Input translator - system executive
2) Process simulation subsystem
3) Costing and evaluation subsystem
4) History and report writers.

An evaluation system executive is a relatively simple sub-program in that its only task is to direct tanslated input information to the simulation subsystem, the evaluation subsystem and history writer. In some cases though this stage can be a little more complex if the executive acts as a preprocessor stage and determines the size and type of evaluation to be carried out, collects those subroutines needed and writes a calling program to carry out the evaluation; in this way only those subroutes needed are loaded, thus saving computer storage albeit at the cost of longer running times. This preprocessor approach is becoming more common in the large process engineering systems found in the chemical engineering industries.

The problem of input style, especially with respect to engineers unfamiliar with computing, has long since been recognized in industry and since there is a variety of input subsystems available, this is no longer so much a problem area as a matter of tailoring a subsystem to meet the particular application; certainly, especially with the widespread use of problem-oriented languages and interactive terminals, input systems are now available for those with even only a very rudimentary knowledge of computing. Also the continuing decline in costs of graphical devices and the broadening availability of graphical software has made computer graphics a feasible tool in flowsheeting, presentations and analysis (426). It should however be remembered that, whilst easily used input formats and "cosmetic" presentation are very important in industrial applications and that systems should be designed with this end use in mind, development of the underlying systems technology is initially likely to be of considerably more importance.

Although systems for equipment sizing and costing (24) and process evaluation (25) exist for the minerals industry, these would require a lot of adaptation to be of use in an integrated evaluation system. Thus it would be expected that if such a system were to be developed from scratch the costing and economic evaluation section should also be developed specially, although of course it could use parts of pre-developed packages. The task of the subsystem executive would again be fairly straightforward in that it need only accept and direct

information and call calculation subroutines in a certain predetermined manner. Although in the design procedure, equipment sizing and capital costing may be calculated separately, this would add another calculation stage and degree of complexity and so these two phases would be integrated in the proposed system. Information for the equipment sizing calculations would be obtained from the history file containing the process simulation results and/or from the input translation step. The data retrieval should be written in such a manner that system determined parameters may be overidden by the user; in this way, not only can the experience of the engineer be used to enhance any system weaknesses but also the subsystem is likely to be of use in a "stand-alone" fashion without depending on the results of a process simulation. The corollary of this is that the simulation system should also be designed to be used alone. In the same way that the simulation uses unit operation modules to calculate mass and energy balances, so the evaulation calculations should use similar modules for equipment sizing and costing. Operating costs can also be calculated from these modules and also from the simulation results, as can projected revenues; from these, and using taxation modules, the executive can determine project cash flows and in turn use these in an economic analysis. The results from these stages would then be sent to the calculation history file from which a detailed report can be written.

Once the various stages of calculation have been completed the system would need to produce a report of the results. This could be compiled from the calculation history file which stores results of calculations as they proceed, in case a correct solution is not found, the final results being used in the report writing stage. This stage is likely to be the simplest in the whole system, the only complexity likely to arise being the possible, and useful, requirement that the user can specify in the input the amount of detail presented in the final report. Some users may only want a clear and concise summary of the process evaluation whilst others would want details of the behaviour of each component in the process. However, just as with the input stage, it should be remembered that such facilities are "cosmetic" and should not sidetrack development effort away from the underlying sub-systems such as simulation and evaluation.

As can be seen from the above the development of a system for the

evaluation of metallurgical processes should not be too arduous a task once the two main constituent sub-systems, simulation and evaluation, are developed. The problems involved with the latter have already been discussed and so successful completion of an evaluation system largely relies on obtaining or developing a method for simulating, i.e. obtaining steady state mass and energy balances around a metallurgical process. The simulation subsystem is the core of any evaluation program and is usually where most development effort is directed.

The use of simulation systems, also known as flowsheeting systems, in the chemical engineering industry has flourished since the early 1960's and their use has now become widespread and indeed commonplace; many systems have developed in-house and often serve as a repository of a company's expertise. Also, in a similar manner to that proposed here for minerals extraction processes, there has been a trend in the chemical engineering industry towards integrating simulation systems into much larger systems for project engineering and evaluation (427). As Ford (16) has shown, the minerals industries lag behind these developments by some 20 years even when only considering process simulation.

The development of one of the more flexible, general purpose simulators involves considerable effort, typically consisting of 20 - 60 man-years, over 3 - 5 years, and a cost of several millions of dollars (428). Only large organizations are capable of managing such development programmes and have the high rate of useage to justify the effort, so most programs have been developed by universities or large engineering and producing companies. Systems from universities have mostly concentrated on the demonstration of new solution methods or are used for teaching purposes; generally they are fairly basic in format and facilities and as packages are of relatively little industrial significance. However due to the nature of academia, this work is usually detailed in the literature and are fairly easy to obtain and use. In the commercial environment, more sophisticated, robust and user-oriented systems are produced and used, but these are normally not described in the literature and are both difficult and expensive to acquire (427, 429); at the time of writing the normal rate for leasing a commercial evaluation system such as DESIGN/2000 (430), PROCESS (431) or CONCEPT (432) was around $U.S.50,000 per annum (433). It should be noted in this context though, that in recent years some companies, as

they have developed new systems, have made their now obsolete older programs available for wider use; thus FLOWTRAN (434) and FLOWPACK (435), large general purpose packages developed by Monsanto and ICI respectively, have been installed on university systems both in this country and in America (436, 437). Another encouraging note towards more readily available systems comes from the U.S. Department of Energy funding of the development of a major new system, ASPEN (Advanced System for Process ENgineering) (438, 439), at Massachussetts Institute of Technology; a large government grant enables a 'state-of-the-art' process evaluation package to be produced for the analysis of coal liquefaction and gasification processes. Under American laws, the results of such a government sponsored project must be made publicly available at a nominal cost; release of this public version of ASPEN was expected at the end of 1982 at a probable cost of $1000 (433). Also an improved version, ASPEN PLUS, developed privately by the authors of the original program, is to be released, under control, to a limited number of academic institutions for further development (433, 440).

Despite the limited number, quality and availability of systems in the literature, it was decided to attempt to base the methodology on existing simulation software, if necessary by drastic adaptation. In coming to this conclusion, the long lead times and high development costs for similar systems were taken into consideration, as was the likelihood that third-generation chemical engineering systems, already designed, would soon be able to handle solids processes. It was also hoped that the limited application of fairly simple and robust structures for process simulation proposed by King (441) and Ford (16), which are restricted to physical separation and transformation processes, could be improved upon.

4.2 Mineral Processing Simulation

At the time (Summer 1981) that a final decision had to be made on what software would be used to illustrate the methodology, a survey of the literature on minerals extraction plant simulation was, by nature, very brief. Whilst a number of authors (12-15) had reviewed the state-of-the-art and it was claimed (12, 13) that the use of simulation in design was common and would become widespread within a decade, there was no existing general purpose simulator suitable for the methodology

and its proposed applications.

Two simple balancing programs (17, 18) have been published but these use general models such as component separators relying on a priori knowledge of the separation characteristics and are too basic to allow streams to be described in any detail. Similar restrictions beset the only program (5) claiming to simulate mineral processing plants in any detail; the calculation procedure relies on linearized models of unit operations thus precluding the commonly used phenomenological models intended for the methodology.

The only published successful application of process plant simulation in the minerals industries was a program designed to simulate coal preparation plants (129). This however used a fixed stream structure describing particle size, density and composition and requiring water to be the only fluid component. Also the calculation procedure restricted the number of recycle streams to three and could not guarantee solution of these.

The structure for ore dressing plant simulators proposed by King (441) and Ford (16), whilst offering some progress, is similarly restrictive in only allowing water as a fluid component and not permitting chemical separations such as leaching. This structure has been used by simple flotation circuit simulators (218, 236, 237) but, whilst a large development in Canada (39) uses a similar technique, no successful general purpose simulator has yet been published using it.

Thus the most likely route to rapid progress appeared to be through the adaptation of a chemical process plant simulation program. It was considered that, especially with the application of FLOWTRAN to coal pretreatment processes (23), this was a feasible and logical approach although use of an already developed and suitable system would be preferable (442). Thus, with this development in mind and an appreciation of the differences between mineral and chemical process simulation requirements, the state-of-the-art in chemical engineering simulation systems was studied.

Since that decision was made descriptions of three simulation systems, in development and/or specifically intended for use in minerals extraction processes, have been published. FLEXMET (11, 19) and APEX (20), written for the Fluor and AMAX companies respectively, are two industrially developed simulation systems available only in-house or at commercial rates of about $50,000 per annum (433); whilst broad

descriptions have been published in the literature, neither company appears willing to discuss the details of program structures and procedures. More recently, the University of California, Berkeley, has described the design aspects of its UCMINPRO system (21), which is still under development.

Although few details of these systems are available, some initial conclusions can be drawn from the published descriptions:-

i) minerals companies are beginning to realize the potential usefulness of and benefits likely to accrue from the development of minerals extraction plant simulators and are prepared to invest accordingly.

ii) whilst techniques developed in the chemical industry are being utilized, software is not and so a great deal of effort is being expended re-inventing the wheel; it would appear that the calculation procedures used, whilst fairly robust, are relatively inflexible and lack the sophistication developed over 20 years by the chemical industry.

iii) due to the excessive, and perhaps misguided, effort required to develop the executive systems, it would appear that little work has been carried out on the unit operations modelling aspect of minerals extraction process simulation and no efforts have been made to integrate the simulation with process evaluation routines; use of the APEX and FLEXMET systems is limited due to the use of simplistic mathematical abstractions rather than engineering models to describe the unit operations and the need for a priori knowledge of streams and separations for some processes.

Therefore, whilst it is encouraging that the minerals industry is recognizing the possible value of computer-aided approach to plant design and analysis and that progress is being made, it would appear that some of the effort is perhaps wasted and the area where most advances need to be made, unit operation modelling, is still relatively ignored.

4.3 Differences between Chemical and Mineral Process Simulators

As Evans and Seider (4) have pointed out there are a number of major limitations to using a 'second-generation' chemical process engineering, computer-aided design system; the problem areas are:-

1) Unit operation subroutines for solids handling processes

2) Physical property data and estimation methods

3) Inability to handle solid-liquid and solid-vapour streams

As has been shown in Chapter 2, the first of these need not be a major obstacle to the development of mineral extraction systems. There are enough models reasonably well developed to illustrate the wide application and potential value of such a system.

The second problem area, that of physical property determination, is also not a major obstacle. Due to the nature of chemical processes, often operating with large variations in temperature and pressure, accurate and fast estimations of physical properties is essential in plant simulations; in some packages, the physical property estimation and data routines comprise almost 50% of the total. Many minerals extraction plants operate near ambient conditions and so the need for property prediction is considerably reduced; even in those plants where this capability would be useful, notably hydro- and pyrometallurgical, the number of species of interest would be relatively small. This is especially so when it is considered that few solids, particularly mineral, properties can be predicted from data banks and that metallurgical parameters such as grindability and floatability invariably have to be determined from testwork for any new ore. Indeed, parameter determination is potentially the major drawback to widespread computer-aided design techniques, in that the value of the technique is reduced as the required amount of input data increases.

Therefore the large and sophisticated property packages that form an integral part of most chemical engineering simulation systems will be largely redundant in metallurgical applications; if, however, hydrometallurgical and pyrometallurgical systems are to be studied then a similar, but considerably smaller, package would be needed for the compounds of interest, such as water, air, inorganic solvents etc. Due to the high modularity of the systems it is anticipated that this substitution would cause few problems and, along with the replacement of inappropriate unit operation models, should drastically reduce the computer storage needed.

The most important limitation in Evans and Seider's list of problem areas is the inability of chemical process simulators to permit adequate representation of the properties of solids; provision must be

made to represent solids' variables such as particle size, shape, specific gravity, floatability and mineral composition. This problem arises from their use of dimensioned arrays characterized by the numbers of components and their variables and the number of streams. The programs therefore tend to be inflexible, as the program designer must know all variables of interest at the time the program is designed. It becomes difficult to incorporate new types of variables without altering the layout of data and changing every routine that was defined on the basis of the original layout (23). Even if a program could handle solids, if a process plant to be simulated consisted of a number of process units and several particle classes this type of formulation would be computationally inconvenient and inefficient (441). For instance a crushing plant needs a particle size distribution describing very coarse to very fine particles; it may then be followed by a specific gravity pre-concentration stage which would be described using a finer particle size range and a specific gravity distribution; the subsequent grinding and flotation circuit simulation needs yet finer particle sizes, grade distribution and possibly also flotation distribution. Thus any simulation system capable of modelling a diverse range of processes must be based on a flexible stream structure.

An alternative type of data structure has been suggested (22) based on a plex rather than on fixed arrays. In a plex structure, information is stored in blocks of contiguous storage locations known as beads; these beads, of any length, are created dynamically from a pool of free storage as needed during execution of the program and are referenced and linked together by means of pointers. Not only are they used for describing stream structures, but they can also be used for all other data handling, including process unit variables and plant flowsheet descriptions; in this manner, the only restrictions placed on a plant simulation would be physical and caused by the size of the computer used. Example beads and their uses are shown in Fig. A1-2. A prototype system, PLEXYS, was implemented to study the use of plex data structures, and integrated with FLOWTRAN to illustrate their potential to model solids handling equipment (23). The first system designed specifically to use the plex structure was ASPEN (438, 439) and its use in describing different stream structures is described in Chapter 5 and Appendix 1. It is believed that, to date, ASPEN and its

successor, ASPEN PLUS, are the only chemical engineering systems, thoroughly tested and in commercial use, that are capable of handling a variety of solid stream structures.

4.4 Chemical Process Engineering Evaluation Systems

Literature on the use of chemical process simulation as a design tool began appearing in the late 1950's (443, 444) and a general purpose system was produced as early as 1960 (423). From initial rudimentary mass and heat balancing algorithms, the growth of "flowsheeting" ("the use of computer aids to perform steady-state heat and mass balancing, sizing and costing calculations for a chemical process" (445)) has had a major impact on the art of chemical process design (427), so that today, no major chemical or petrochemical plant would be built without first simulating the process on a computer by use of mathematical models (428); even in 1974, Villadsen was able to find 74 operational flowsheeting systems (446).

Motard et al (27) described the history and development of computer-aided design until 1975 and the characteristics of the early systems and their limitations were defined by Evans and Seider in 1976 (4); the wide variety available at that time was reported in a series of articles (447, 448) originating from the ASPEN project. Since then further reviews (427, 428, 445, 449, 450) have described further developments and trends in the direction of the field. Those developments relevant to the production of a flowsheeting system and methodology for minerals extraction plants have been in the area of system architectures and their associated data structures and calculation algorithms.

A flexible computer-aided evaluation system will contain a fairly large library of unit operation models, property evaluation routines and cost estimation routines; such a system enables an engineer to develop and solve an arbitrarily configured process model with relative ease. However this model could consist of many thousands of simultaneous non-linear equations and, whilst Sargent (451) has reviewed generalized methods, specialized approaches for solving such models have evolved. On the more common approaches, 3 groups of system architectures have been based; these are the sequential modular, equation-oriented and simultaneous modular structures.

Sequential Modular Systems

Almost all industrial flowsheeting systems have been based on the sequential modular approach. In these a module is developed for each unit operation in a flowsheet to calculate the outlet stream variables as functions of the inlet stream variables and process parameters; these modules are called in turn to simulate a process, the streams acting to carry information through a simulation in analogous fashion to the transport of mass and energy in the real process. As has already been discussed, it is only recently that the stream structures capable of carrying information to describe a solid-liquid stream have been developed thus eliminating the major drawback of this approach.

Recent improvements to systems using this approach have centred around the ordering of module calculations when handling recycle loops and the acceleration of the convergence of these calculations to a solution. To solve a process flowsheet by the sequential modular it is necessary to partition the flowsheet, select tear streams, nest the computations and determine the computation sequence. Whilst engineers can often perform these steps by inspection, for relatively simple flowsheets, the task can become arduous and difficult for more complex problems. In older systems the program user must supply the computational sequence as input; however algorithms for automation of this have been developed although, whilst useful in most cases, there are situations when the optimal sequence is not obtained and so a manual overide capability is needed.

More efficient algorithms for convergence of the approach to a solution have also appeared in the literature recently. However none of these are the best method for all problems and so recently introduced flowsheeting systems offer the user a choice.

Using the modulesin sequential modular systems, one cannot easily specify an output specification, i.e. it is difficult to use them in a design mode. As this has been one of the architectures major limitations a lot of work has been carried out to alleviate the problem. Design specifications can now be handled by placing a control block around a module and adjusting a parameter within that module such that the desired specification is met. Whilst this technique does have solution stability, it is generally most efficient when operating on only one unit operation

and can add considerably to the run time of a simulation.

Equation - Oriented Systems

In order to circumvent the inefficiencies associated with the design specification problem for sequential modular systems, there has been considerable academic effort to investigate how to perform all computations simultaneously. Prototype systems such as GENDER (452) and SPEED-UP (453, 454) have been developed in universities but a thoroughly tested and easily useable system has yet to be introduced.

The approach offers many advantages over more traditional methods such as speed and possible optimization; in the proposed methodology, the problem of describing numerous stream structures disappears as streams are no longer used to carry information. However acceptance of the approach has been slow because of the complexity of the executive routines and numerical calculations involved and the very large computer storage needed for solutions; the user interface is also problematical in that if a solution fails, few diagnostic messages can be written and introduction of new routines is not as well defined a task as other architectures.

The SPEED-UP system has been in almost continuous development for over 20 years (453-455) at Imperial College and a dependable version has yet to be produced; a preliminary edition is presently available but has to undergo thorough testing before widespread release (455). However when ready for use this type of system is likely to prove a serious competitor to the sequential modular architecture.

Simultaneous Modular Systems

In order to retain many of the advantages of the equation solving approach, and much of the software of the sequential modular systems, but overcoming some of their disadvantages, an architecture known as simultaneous modular has been mooted by Rosen (456). Rigorous unit operation modules, usually those developed for sequential module systems, are used to determine parameters for simple linear models which are represented as equations. These are then solved by an equation solving technique to estimate stream variables which allow the rigorous models to be called again. This technique is intended to be faster, especially for complex flowsheets, and should also allow more design specifications and optimization approaches. Whilst 'third generation' programs such as ASPEN (438) and FLOWPACK II (457) are designed to take advantage of

this architecture, no system has yet been introduced designed around it, as its possible uses and problems have yet to be fully explored.

4.5 Summary

If the methodology is to be exploited to its full potential, a powerful and flexible executive system is required to handle the complex calculations involved with mass and energy balances, equipment costing and economic evaulation. The development of such a system can take up to 60 man-years and so it is logical to utilize existing software. However most useful software has been written for the chemical process industry and is based on inflexible data structures unsuitable for minerals extraction processes. A solution to this problem, the plex structure, has been proposed and can be used to adapt existing software. Most useful systems have been developed industrially and are difficult or expensive to obtain. However the ASPEN project has written a stage-of-the-art flowsheeting system based on the plex structure and the software can be acquired, especially for academic purposes. As it is the only system using this structure it is therefore the logical choice for basing the methodology around; a description of the ASPEN PLUS system is to be found in Appendix 1.

At present this type of system can only be used for simulation purposes and so further workers in this area should remain aware of different approaches to solution calculation as they are developed. The SPEED-UP system should soon be available and will allow design and optimization calculations to be made more readily.

CHAPTER 5 Implementation of the Methodology

5.1 Process Mass and Energy Balance Calculation

5.1.1 Introduction

As has been described in the previous chapter, a computer program used to calculate the mass and energy balance of a process can be also used as the core of an overall process evaluation system modelling both the operating and the economic aspects of the process; the development of such a system is very time-consuming and so it is logical to utilize existing software wherever possible. As was also discussed, the most suitable system existing appears to be the ASPEN PLUS system; this is based on computer software developed for the U.S. Department of Energy to study synthetic fuels processes, which has now been debugged, developed and is maintained by a company consisting of many of the development team of the original ASPEN system. A description of the ASPEN PLUS system is to be found in Appendix 1. The use of this system at the company's site in Cambridge, Massachussetts, was kindly offered to the author and also, eventually, a copy was provided for it to be implemented at the University of London Computer Centre for future use and development of the methodology.

As is described in Appendix 1, the ASPEN PLUS system is based on a flexible data structure and modular system architecture. Whilst offering advantages, the major importance of the data structure with respect to the modelling of minerals extraction processes is its ability to handle stream structures of almost any form; whilst based on a certain, defined structure it is possible to add any number of additional descriptive data sets and therefore to characterize almost any form of process stream. The combination of such a flexible stream structure with a completely modular system architecture meant that to adapt ASPEN PLUS to simulate extraction plants, it was necessary only to add the desired classes and suitable unit operation models.

5.1.2 ASPEN PLUS Stream Classes

Whilst ASPEN was originally developed to simulate processes treating solid raw materials such as coal and oil shale, there is no suitable stream class for simulating the more common minerals extraction processes. Solids are generally treated as non-conventional components[1/] whose

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[1/] For explanation of terms such as components, substreams, attributes, etc., refer to Appendix 1, where such terminology is explained.

properties may be described using component attributes and sub-stream attributes. Component attributes are used to define the characteristic physical properties, such as density, of each non-conventional component in a simulation, whilst sub-stream attributes are used to describe the overall properties of a substream. Although the use of component attributes is applicable to the simulation of minerals extraction processes, it was found to be necessary to add extra sub-stream attributes to ease the use of the system for such simulations. The only existing sub-stream attribute describes the particle size distribution; it would be possible to describe a heterogeneous mineral population using a number of sub-streams, each with its own particle size distribution and describing one mineral class. However not only would this be clumsy and inefficient to use if there was to be a number of such mineral classes, but it would also have proven difficult to allow for the liberation of minerals if a suitable unit operation model of comminution was implemented.

Therefore a substream attribute, named PSMFD, was implemented on the test version of ASPEN PLUS used during the project development stages. This substream attribute may be used to describe a solids substream divided into a number of particle size classes, these being further sub-divided into a number of mineral fraction classes. Each substream is divided into valuable mineral components and gangue mineral components, the mineral fraction of a particle being defined as the fraction of the particle consisting of valuable minerals; any number of valuable and gangue minerals may be present, but the mineral fraction considers all valuable minerals as the same and all gangue minerals similarly. An attribute may contain an arbitrary number of particle size and mineral fraction classes each with arbitrary limits; a stream can also consist of any number of substreams described by such an attribute, though the properties and classes of each may be different. This flexibility highlights the suitability of the ASPEN PLUS stream structure for use in the simulation of minerals extraction systems. Figure 5.1 (a-c) shows how a typical minerals stream may be defined using the PSMFD substream attribute, the ASPEN PLUS stream structure and their associated input language; the only difficulty in using the attribute is that it is necessary to ensure that the individual component flows relate to the fraction of the total flow as defined by the substream attribute.

```
COMPONENTS METAL/GANGUE/H2O H2O
ATTR-COMPS METAL  CAUSR1 / GANGUE CAUSR2
DEF-SUBS-ATTR PSMFD PSMFD
  INTERVALS NPSD=5  NMFD=5 NTOT=25
  SIZE-LIMITS 0.000290/0.000220/0.000160/0.000110/0.000075/0.0
  MIN-FRAC-LIM 0.7/0.5/0.38/0.25/0.1/0.0
DEF-STREAM-CLASS MNCPSMFD
  DEFINITION SUBSTREAMS=MIXED NCPSMFD
DEF-STREAMS MNCPSMFD ALL
DEF-SUBS-CLASS NCPSMFD
  DEFINITION TYPE=NC ATTRIBUTES=PSMFD
DEF-SUBS NCPSMFD NCPSMFD
STREAM S1
  SUBSTREAM MIXED MASS-FLOW=180.0 TEMP=300.0 PRES=1.0 &
                FLASH-OPTION=NOFLASH
                   MASS-FRAC H2O 1DO
  SUBSTREAM NCPSMFD MASS-FLOW=120.0                    &
              TEMP=300.0 PRES=1.0
  MASS-FLOW METAL 43.77 / GANGUE 76.23
  COMP-ATTR METAL CAUSR1(7860.0)
  COMP-ATTR GANGUE CAUSR2(2700.0)
  SUBS-ATTR PSMFD FRAC=0.045 0.03 0.03 0.0225 0.0225          &
                      0.045 0.03 0.03 0.0225 0.0225          &
                      0.060 0.04 0.04 0.0300 0.0300          &
                      0.060 0.04 0.04 0.0300 0.0300          &
                      0.090 0.06 0.06 0.0450 0.0450
```

Figure 5.1(a) ASPEN PLUS input language to describe a minerals stream

100.0	Water flow Sstr MIXED	0.045	Sstr. Attr.
100.0	Total flow	0.03	PSMFD
300.0	Temperature	0.03	
1.0	Pressure	0.0225	
0.0	Enthalpy	0.0225	
0.0	Molar vapour fraction	0.045	
0.0	Entropy	0.03	
1000.0	Substream density	0.03	
18.0	Substream mol. wt.	0.0225	
43.77	Metal flow Sstr NCPSMFD	0.0225	
76.23	Gangue flow	0.060	
120.0	Total flow	0.04	
300.0	Temperature	0.04	
1.0	Pressure	0.0300	
0.0	Enthalpy	0.0300	
0.0	Molar vapour fraction	0.060	
0.0	Entropy	0.04	
1.0	Substream density	0.04	
1.0	Substream mol.wt.	0.0300	
7860.0	Metal density CAUSR1	0.0300	
⋮		0.090	
2700.0	Gangue density CAUSR2	0.06	
⋮		0.06	
		0.045	
		0.045	

Figure 5.1(b) Stream data vector set up by input shown in Figure 5.1(a) (descriptive data not shown)

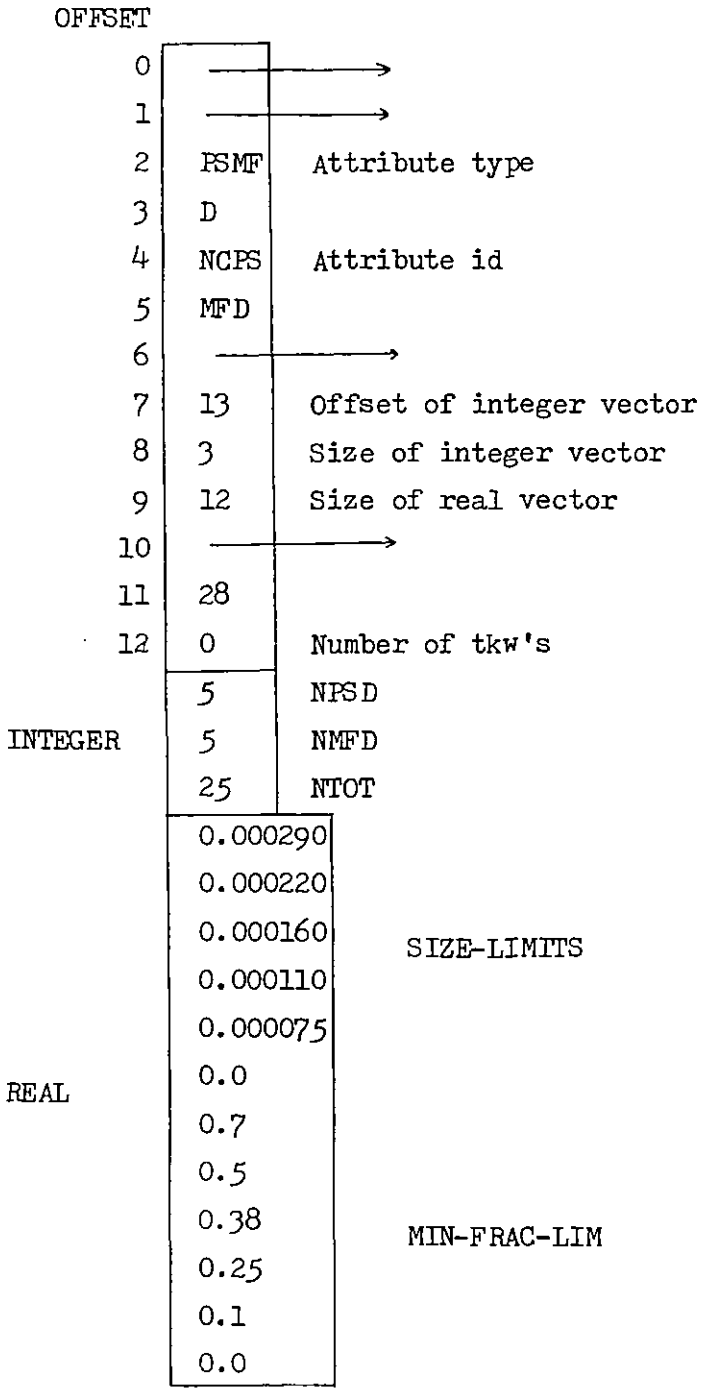


Figure 5.1(c) PSMFD substream attribute structure as set up by input language shown in Figure 5.1(a)

To implement and use the new substream attribute it was necessary, not only to insert a new System Definition File table defining the new input language and plex structure, but also to develop utility subroutines to handle the data structure and incorporate these into unit operation modules to enable them to process the new substream types. As is described in Appendix 1, the purpose of the unit operation interface subroutine is to access the data storage locations in the plex and manipulate these into a more manageable form for use in the unit operation model subroutine; generally the interface only handles the location of each data storage set and calls utility subroutines to actually carry out the manipulations. General purpose utility subroutines were written to access the PSMFD attribute plex locations and carry out the following manipulations:-

(i) store the size and mineral fraction class limits of each attribute in separate arrays in descending order (subroutine STPSMF)

(ii) develop a two dimensional array for each substream carrying the mass flowrate in each size and mineral fraction class in descending order; this is as opposed to the plex data structure in which the mass fraction in each class is stored in a single list (subroutine STPSMF)

(iii) following unit operation model calculations, transfer the flow data from arrays into the appropriate plex storage locations (subroutine STPSMF)

(iv) change data in arrays describing the flow in each size and mineral fraction class into a one dimensional array describing only the flow in each size class (subroutine PSMFPS)

(v) carry out the manipulation in (iv) in reverse, assuming that the mineral distribution in each size class is the same as before the model calculations (subroutine PSPSMF)

(vi) calculate the result of mixing a number of subroutines with the same PSMFD attribute (in ASPEN PLUS module SSAMIX)

(vii) write a substream attribute report for each PSMFD attribute in a simulation (subroutine PSMFRP and ASPEN PLUS module ATTRPT)

Having implemented the substream attribute and its supporting utility subroutes, it was necessary to develop unit operation modules which could handle the new attribute and use this new capability to simulate minerals extraction processes. As is discussed in the subsequent sections (5.1.3 - 5.1.11), a set of new unit operation modules has been implemented on the ASPEN PLUS system to carry out these simulations; all of the new modules which process solids bearing streams (crushing, grinding, sizing, flotation, leaching and solid-liquid separation) are capable of handling substreams containing the new attribute and the new comminution, sizing and solid-liquid separation modules are also capable of handling substreams containing the already existing PSO substream attribute. The general purpose ASPEN PLUS modules for mixing, splitting and class changing can now also handle the PSMFD attribute. A more detailed explanation of how the unit operation module interface handles and manipulates the plex stream data structure is given in Appendix 1.

For future use in the simulation of coal washing and other similar processes, another substream attribute, PSSGD, was implemented to describe the distribution of particles in particle size and density classes. This is similar in structure to the PSMFD attribute, though in this new attribute the ratio of component flows is assumed to be the same in each size and density class. The attribute is therefore likely to be most useful when there is only one component flow, e.g. "coal", in a substream and indeed would become redundant if separation models were to use the PSMFD attribute, calculating particle density on the basis of the amount of each mineral present; however the approach of treating the solids flow as homogeneous, with respect to components, has been found to be useful and valid in coal preparation plant simulators (129), important variables such as stream ash content being calculated from the specific gravity analysis rather than from actual component analysis. A similar set of utility subroutines to those developed for handling the PSMFD attribute, were developed for handling the PSSGD attribute, but the only unit operation modules which are presently capable of using the attribute are the mixing and the new comminution and sizing modules; for full use of the attribute it is still necessary for suitable separation modules to be implemented.

Having implemented these substream attributes it was still necessary

to decide on a method and format for describing other solids properties such as component density, breakage parameters and flotation rate constants. Component density is a property which is independent of other variables and which will remain constant throughout a processing plant; this could therefore be contained in the stream data structure as part of a component attribute data set. Breakage parameters, such as work indices and breakage rate and distribution functions, and flotation rate constants are, however, more variable and are more dependent on other varaibles such as equipment parameters and operating conditions; it was therefore decided that this type of variable should not be carried in the stream data, but that it would be more appropriate for their values to be included in each unit operation module's input data.

5.1.3 Crusher Module

The commercial version of ASPEN PLUS includes a crushing model developed from the U.S. Department of Energy coal preparation plant simulation system (129), which can be used to simulate gyratory jaw, roll and cage crushers. Product size distributions are calculated using Bond's law and built-in matrices for the selection and breakage functions; whilst it is recognized that these functions change between classes of crusher, there is no allowance made for the crushing characteristics of different materials, different machine operating modes and the internal classification which occurs in many crushing machines.

Due to these deficiencies, it was decided to implement a second crusher block which would be more appropriate for the simulation of rock crushers for design purposes. Whilst Whiten's model (92) would be more suitable if an existing operating plant was to be simulated based on the results of testwork to determine model parameters, it was felt that the best model for a design application would be based on Bond's law, Allis-Chalmers' definitions of crusher power rate and empirical relationships describing the crusher product size distribution.

The unit operation block CRUSH2 simulates the performance of a number of crushing machines operated in parallel and is able to model jaw, gyratory and cone crushers, both in open and closed circuit. The model is based on the following assumptions and relationships:-

Crusher Power Rate (79)

The power rate of a crusher is defined as its ability to inject

energy per unit of feed; experimentation has determined that this ability is dependent on the shape and size of the crushing chamber, feed rate and setting control, position in the crushing circuit and installed motor size. It has been observed (79) that the following average crushing efficiencies prevail:-

For primary and secondary crushers:-

% Efficiency = 60 for machines with manual control

% Efficiency = 70 for machines with automatic feed rate control

% Efficiency = 90 for machines with automatic feed rate and setting control

For tertiary and subsequent stages:-

% Efficiency = 80 for machines with manual control

% Efficiency = 90 for machines with automatic feed rate control

% Efficiency = 95 for machines with automatic feed rate and setting control

The crushing efficiency is the percentage of installed power that appears to be used in crushing according to Bond's law when a crusher is operated at its minimum efficient feed tonnage, and, therefore, maximum power rate. These efficiency values are used to define a crusher's effective power rate from the user's definition of installed power, position of crusher in circuit and control type; whilst it would have been more efficient for the user to define a crusher's power rate, few such rates have been defined for the various types and sizes of machines. Especially when a crusher is operating in closed -circuit, it is important to ensure that the actual tonnage passing through is approximately the same as the design tonnage; a future update of the block should allow the user to define the design tonnage, the block then calculating the number of crushers needed to handle the actual tonnage.

Bond's Law (63)

It is necessary for the user to determine, from laboratory testwork, the value of the Bond Work Index of the rock being crushed; whilst Bond specified a test to determine the crusher work index, it has been found more reliable to use the rod mill work index. Whilst Bond's law is a purely empirical relationship, it has been found through extensive industrial use to provide a reasonable prediction of the size reduction achieved in comminution processes; the law can be expressed as:-

$$W = 10W_i \left(\frac{1}{\sqrt{P_{80}}} - \frac{1}{\sqrt{F_{80}}} \right)$$

W = Energy consumption/unit wt.

W_i = Work Index (kWhr/ton)

P_{80}, F_{80} = 80% passing size of crusher product + feed. (μm)

Material Homogeneity

No model developed to date of the crushing process has successfully described the liberation effects caused by crushing; the CRUSH2 unit operation block treats all feed and product material as homogeneous, the total weight of material in a mineral or density class remaining constant. Whilst this may be regarded as inaccurate there is still an inadequate knowledge of the liberation phenomena and mineral distribution is not usually considered important until after the crushing step.

Product Size Relationships

When designing a crushing plant using traditional methods it is normal practice to determine the crusher product size distribution using tables or graphs published by the major equipment manufacturers. As was seen in Chapter 2.1.5, empirical relationships describing the surfaces shown in these graphs have been developed; these relationships, described in Figs. 2.2-2.5, are used by CRUSH2 to predict the complete product size distribution from the product 80% passing size predicted by Bond's law. Whilst these relationships suffer the same dangers and restrictions of empiricism as Bond's law, the data on which they are based has been found in industrial use to be fairly reliable.

A typical set of data input for the CRUSH2 block is shown in Fig. 5.2(a); the input parameters are installed power (PINS), design tonnage (TOND), Bond work index (BWI), crushing stage (MPOS), control type (CTYP), machine type (MTYP), circuit type (CIRC) and number of parallel crushers (NOC). An example of the Input Data and Results sections of the block output is shown in Fig. 5.2(b) and the calculation procedure used in the model subroutine is shown in Fig. 5.3.

5.1.4 Ball Mill Module

Whilst the ASPEN PLUS crushing module models typical coal crushers, there is no unit operation block which simulates the operation of tumbling mills. It was decided to therefore implement a module, GMILL, to simulate ball mills, but that the development of rod and autogeneous mill

```
BLOCK B5 CRUSH2
  PARAM  PINS=373.0 &
         TOND=194.45  &
         BWI=55.54    &
         MPOS=2       &
         CTYP=3       &
         MTYP=3       &
         CIRC=1       &
         NOC=12
```

Figure 5.2(a) CRUSH2 Input language

```
SOLIDSCRUSH2 (CRUSH2):  B5
     INLET = S8                    OUTLET = S9
  PROPERTY OPTION SET SYSOPO

                                ***  INPUT DATA  ***

  INSTALLED KW (PER CRUSHER)  373.00
  DESIGN TONNEAGE               194.4500
  BOND WORK INDEX            55.54000

  POSITION IN CIRCUIT(1Y,2Y...)2
  CONTROL SYSTEM TYPE NO. 3
  CRUSHER TYPE NO.        3
  CIRCUIT TYPE NO.(1=OPEN)1
  NO. OF CRUSHERS         12

                                ***  RESULTS  ***
  DESIGN POWER RATING      1.73
  AVE. KW PER CRUSHER    335.70
  TOTAL KW IN BLOCK     4028.40
  ACTUAL POWER RATE        0.61
  FEED 80% PASSING     0.057989
  PROD. 80% PASSING    0.036341
  ACTUAL PLANT TONNAGE   6635.7597
```

Figure 5.2(b) CRUSH2 Output

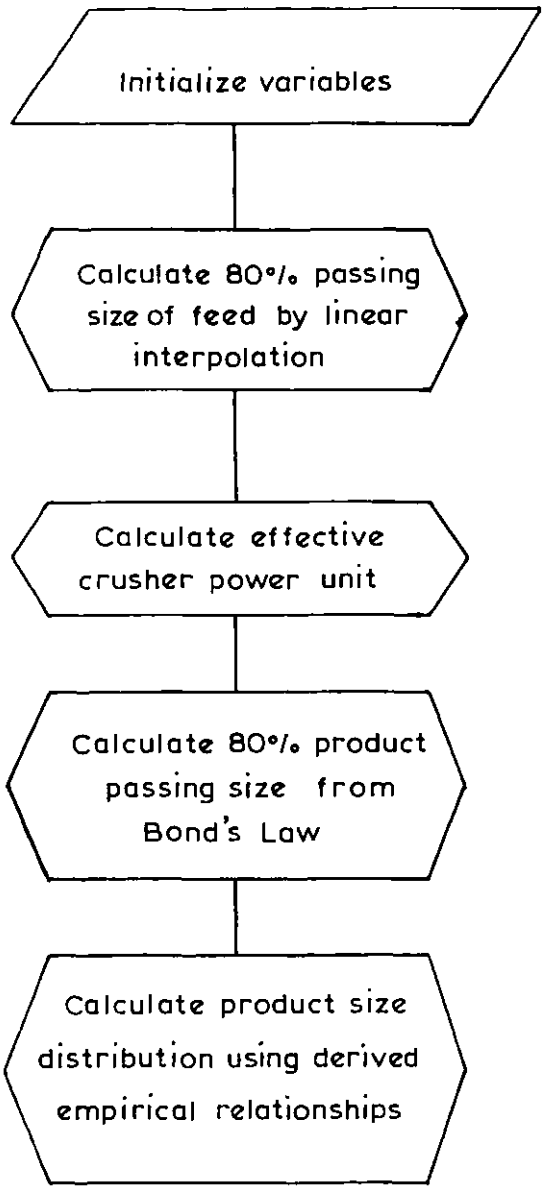


Figure 5.3 CRUSH2 model subroutine calculation procedure

modules would be left to future workers due to the lack of suitable phenomenological models; whilst it would have been a relatively simple matter to model ball and rod mills using Bond's law type correlations, it was felt that the regular use of more detailed, kinetic models for design purposes is likely to become common practice in the near future and replace the use of such correlations.

The unit operation module GMILL implemented on ASPEN PLUS simulates a number of ball mills operated in parallel. The basic way of describing the grinding process, and has evolved from models using probability concepts into a kinetic form. The main relationship used is (14):-

$$P = D.R^{-1} (I - A)^{-1} (F - P)$$

where R = rate at which particles are broken

A = manner in which the fragments of breakage appear in other size ranges

D = rate of particle discharge from the mill

P,F = product and feed size distributions

The actual formulation of this relationship used in the module is based on Herbst and Mika's work (108) which is described and discussed in Section 2.1.6.

It is generally accepted that the breakage distribution, A, is an intrinsic property of an ore, so that its values can be measured in a laboratory scale mill and applied to industrial scale mills with confidence. However the breakage rate function, R, is a measure of the grinding efficiency of a mill for a particular material and so is dependent on the mill characteristics. Due to the lack of reliable scale-up relationships it is necessary to measure this parameter under conditions of complete dynamic similarity with the proposed industrial ball mill. Under these conditions it is believed (109) that the rate of breakage is proportional to the specific power input to the mill, which allows a specific breakage rate function, independent of mill design and operating variables, to be determined and used in scale-up. Whilst this theory is expected to eventually lead to more reliable design procedures, it must be recognized that it is presently largely unproven and that the performance predictions that the module makes should therefore be used with some caution.

The other main variable in the grinding model is the particle residence time distribution of the ball mill. There have been a number

of models proposed to describe this distribution but as there has been little testwork carried out on applying these to industrial mills, the choice of a model is somewhat arbitrary. The model implemented in GMILL is the perfect mixers in series model (115) due to its flexibility, but this is another area where further development work and validation are necessary. It should also be noted that inherent in the scale-up of the breakage rate function based on dynamic similarity, is the assumption that the dimensionless residence time distribution is independent of the size of mill allowing scale-up of the model parameters.

In order to maintain consistency in the use of the scale-up procedure being based on dynamic similarity, the power consumption of each mill is not set by the user, but is instead determined from the relationship:-

$$P = \emptyset.L.D^{2.5}$$

where P = power

$\emptyset$ = a constant independent of mill size

L = mill length

D = mill diameter

The constant $\emptyset$ may be determined from the same tests used to measure the breakage rate and distribution functions. Whilst perhaps restricting the model's flexibility in design work, it is necessary in order to continue the use of the scale-up procedure.

In a similar fashion to the CRUSH2 module, GMILL treats the feed material as homogeneous and does not allow for any mineral liberation.

A typical set of input data for the module is shown in Fig. 5.4; the only variables over which the user has any control in reality are the number of mills (NMIL), mill diameter (DMIL) and mill length (LMIL) though it must be realized that it is necessary to maintain the length to diameter ratio from that used in the laboratory testwork. Other mill parameters which are controlled by the test conditions include the test mill mass hold up (HUPT), test mill power (POWT) and power scale-up constant (PHI1)($\emptyset$). The rest of the input parameters define the material breakage rate and distribution functions, the format of their input utilizing the functional form proposed by Klimpel and Austin (110); this has been found to be applicable to a number of materials and so should not prove restrictive. Its use also has the advantage of reducing the experimentation necessary to determine the model parameters, as well as the data input to the simulation.

```
BLOCK B1 GRINDINGMILL
  PARAM DIAMILL=4.0 &
        MILL-LENGTH=10.0 &
        PHI1=2.18 &
        SF1=0.014 &
        SIZE1=0.0012 &
        ALPHA=0.9 &
        BDF1=0.13 &
        EPSILON=0.3 &
        GAMMA=0.87 &
        BE=3.5 &
        TEST-HOLDUP=150000.0 &
        TEST-POWER=650.0 &
        RTAV=600.0 &
        NMIX=2 &
        NMIL=1
```

where:-

Breakage Rate Function = $SF1.(xi/SIZE1)^{ALPHA}$

Breakage Distrubution

Function = $PHIJ(xi-1/xi)^{GAMMA} + (1-PHIJ)(xi-1/xj)^{BE}$

and $PHIJ = BDF1(xi/SIZE1)^{-EPSLON}$

xi = particle size of the ith interval

Figure 5.4 GMILL Input Language

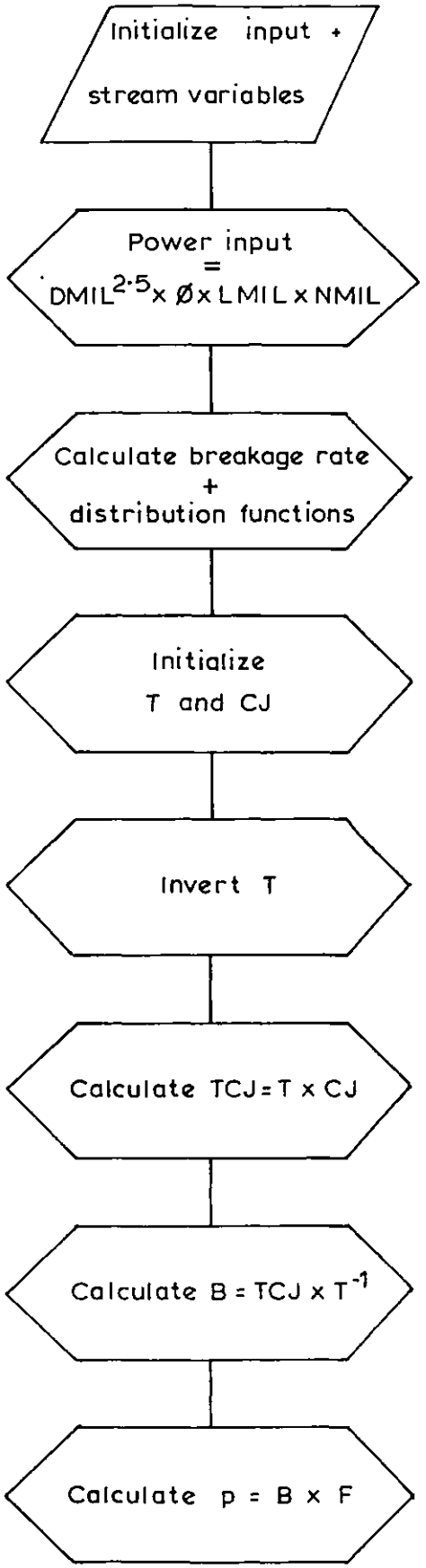


Figure 5.5 Calculation procedure used in GMILL model subroutine

As was stated above the calculation procedure used in this model subroutine is based on Herbst and Mika's formulation (108) and is illustrated in Fig. 5.5; effectively the routine, after calculating the breakage rate and distribution functions, initializes two matrices or arrays, T and CJ, inverts the matrix T using an ASPEN PLUS mathematical utility routine, and carries out the matrix manipulation:-

$$P = T.CJ.T^{-1}.F$$

where T and CJ are matrices calculated from the breakage rate and distribution functions and the mill residence time distribution

and P and F describe the mass flowrates in each size fraction.

In the unit operation section of the simulation report, the only module output other than descriptive data, material and energy balance information and input data is the calculated power consumption.

5.1.5 Vibrating Screen Module

The screen model used in the commercial ASPEN PLUS system is of similar origins to the original crusher module, being based on the model used in Gottfried's coal preparation plant simulator (129), and is also similar in its unsuitability for a mineral process plant design application. This is because the model relies on the use of a "screen separation strength" constant to calculate the mass balance of particles around the screen; this constant depends on the screen type, screen operating and production rate but is independent of the feed's size distribution. This would appear to be contrary to traditional screen design and the experimental data accumulated by screen manufacturers; whilst this approach could potentially be of use in the future, it is felt that at present insufficient data has been published verifying the model's application and describing how the constant may be determined for a particular screen. Therefore a second screen module was implemented to augment the original with a model which was more closely related to accepted screen design technology.

The model on which this new module, SCREEN2, is based, is Karra's recently published model, which is in turn based on screen testwork and design data accumulated by Rexnord (130, 131). Whilst it is recognized that this is an empirical model, it is one which is closely related to

normal design methods and it is apparent from the discussion in Section 2.2.1 that the more theoretically based models are in need of considerable development before becoming accepted industrially. The equations used to describe and adjust the screen partition curve are described in Table 5.1. Other than the validity of these empirical relationships, the only assumption made is that all fluids report to the underflow stream.

As can be seen from Fig. 5.6 which shows the form of the input for the SCREEN2 module, the only screen parameters which the user needs to define are the square mesh apertures, screen area, wire diameter, angle of inclination and deck type (top or bottom); it is also necessary to define the bulk density of the material passing over the screen. The other model parameters shown in Table 5.1 are dependent on the inlet stream properties and are calculated using the procedure shown in Fig. 5.7. The only results printed in the unit operation report section describe the D_{50} calculation and the undersize tonnage flow.

5.1.6 Hydrocyclone Module

The unit operation module HYCYC2 simulates the operation of a hydrocyclone or a number of hydrocyclones in parallel; it is designed to be used in place of the existing ASPEN PLUS module, HYCYC, which is based on Bradley's model for small size hydrocyclones and dilute slurries, and which is also only appropriate for homogeneous solids.

The calculation procedure is a slightly simplified version of Plitt's model (138) which can be used for design purposes without the definition of model parameters other than hydrocyclone dimensions and the pressure drop across it. It is assumed that enough hydrocyclones are used under those conditions for the feed stream flowrate; it is also assumed that the liquid in the feed stream is water, that water recovery to the underflow is the same as the overall pulp recovery and that the hydrocyclone is geometrically similar to that shown below:-

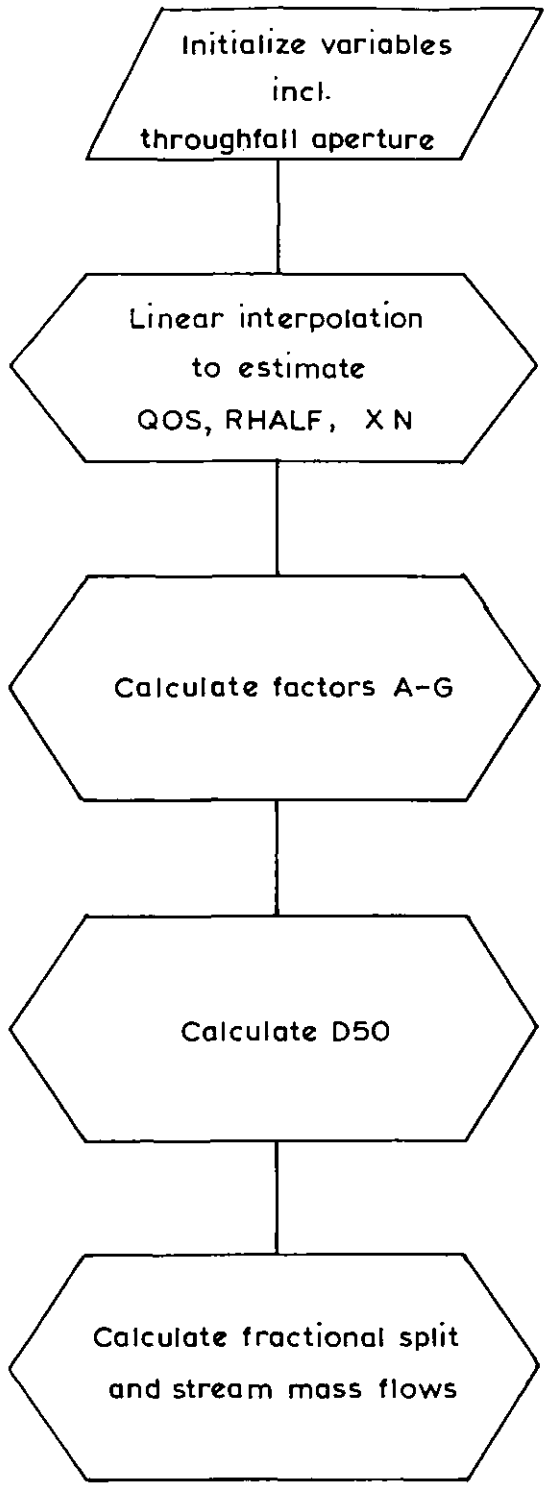


Figure 5.7 SCREEN2 module model routine calculation procedure

```
BLOCK B6 SCREEN2
  PARAM SQ-MESH-APER=0.020 &
        DWIR=0.006 &
        SCREEN-AREA=360.0 &
        THETA=0.349 &
        SCREEN-TYPE=1.0 &
        BULK-DENSITY=1602.0
```

Figure 5.6 SCREEN2 Input language

Table 5.1 : Relationships used to model vibrating screen performance (from Karra (130)).

h_T = screen throughfall aperture (mm) d = wire diameter (mm)
h = screen square mesh aperature (mm) θ = inclination angle to horizontal
d_i = diameter particles in ith size fraction

$$h_T = (h + d)\cos\theta - d$$

$$d_{so} = h_T \left[\frac{\text{Theoretical Undersize (thp/sq.m)}}{ABCDEFG}\right]^{-0.148}$$

A	=	$12.1286\ (h_T)^{0.3162} - 10.2991$	$h_T < 50.8$
A	=	$0.3388\ (h_T) + 14.4122$	$h_T > 50.8$
QOS	=	% oversize in feed to deck	
B	=	-0.012 QOS + 1.6	QOS ≤ 87.0
B	=	0.0425 QOS + 4.275	QOS > 87.0
R	=	% oversize in feed to deck	
C	=	0.012.RHALF +0.7	RHALF ≤ 30.0
C	=	$0.1528\ RHALF^{0564}$	30.0 < RHALF < 55.0
C	=	$0.0061\ RHALF^{1.37}$	55.0 ≤ RHALF < 80.0
C	=	0.05 RHALF −1.5	80.0 ≤ RHALF
S	=	Deck Location : top; S=1; second, S=2; etc.	
D	=	1.1 − 0.1 S	
T	=	$1.26\ h_T$	
E	=	1.0	T < 1.0
E	=	T	1.0 ≤ T ≤ 2.0
E	=	1.5 + 0.25 T	2.0 < T < 4.0
E	=	2.5	4.0 ≤ T ≤ 6.0
E	=	3.25 − 0.125 T	6.0 < T ≤ 10.0
E	=	4.5 − 0.15 T	10.0 < T < 12.0
E	=	2.1 − 0.05 T	12.0 ≤ T ≤ 16.0
E	=	1.5 − 0.0125 T	16.0 < T < 24.0
E	=	1.35 − 0.00625 T	24.0 ≤ T ≤ 32.0
E	=	1.15	32.0 < T
U	=	bulk density (kg/m^3)	
F	=	u/1602.0	
XN	=	% near mesh in feed to deck	
G	=	$0.844\ \ 1.0 - \frac{XN}{100.0}$	

$\%C_i$ = oversize partition coefficient for ith size fraction

$$= 100.0\ [1.0 - \exp(-0.693\ (d_i/d_{so})^{5.846})]$$

The model is based on the following equations:-

$$S = \frac{0.323\ D_c^{-0.39}\ h^{0.54}\ \exp(0.0054\emptyset)}{H^{0.24}}$$

$$R_v = R_F = \frac{S}{S + 1}$$

$$D_{50(c)} = 35.03\ D_c^{0.79}\ h^{-0.45}\ P^{-0.252}\ (\ell_s - 1)^{-0.5}\ \exp(0.063\emptyset)/\exp(0.0055\emptyset)$$

$$Y^1 = 1 - \exp\left(0.693\ \left(\frac{d}{D_{50(c)}}\right)^m\right)$$

$$m = 2.94\ D_c^{0.044}\ h^{0.127}\ P^{-0.084}\ \exp(-1.58R_v)$$

where D_c = internal diameter of the hydrocyclone feed chamber (cm)

h = free vortex height (cm)

$\emptyset$ = volumetric fraction of solids in feed

H = pressure drop across hydrocyclone (as head of feed slurry)
$= P\ \ell_p$

P = pressure drop across hydrocyclone (kilo pascals)

ℓ_p = pulp density (g/cm^3)

ℓ_s = solids density (g/cm^3)

R_v = pulp recovery to underflow

R_F = water recovery to underflow

$D_{50(c)}$ = corrected cut size (μm)

d = particle size (μm)

and Y^1 describes the shape of the corrected partition curve.

These equations are used as shown in Fig. 5.8 to calculate a mass balance. The calculation procedure assumes that a PSSGD type substream attribute is present; if a PSD or PSMFD substream attribute is present the interface calculates the density of each mineral class from component properties. In this way it is possible to allow for the differences in mineral distribution as well as size distribution and so obtain a more reliable prediction of the cyclone performance. The only user input necessary is the hydrocyclone height and diameter (D_c and h) and the pressure drop across it; this simplicity has been made possible by the assumptions made about the cyclone geometry. Similarly the assumption that the

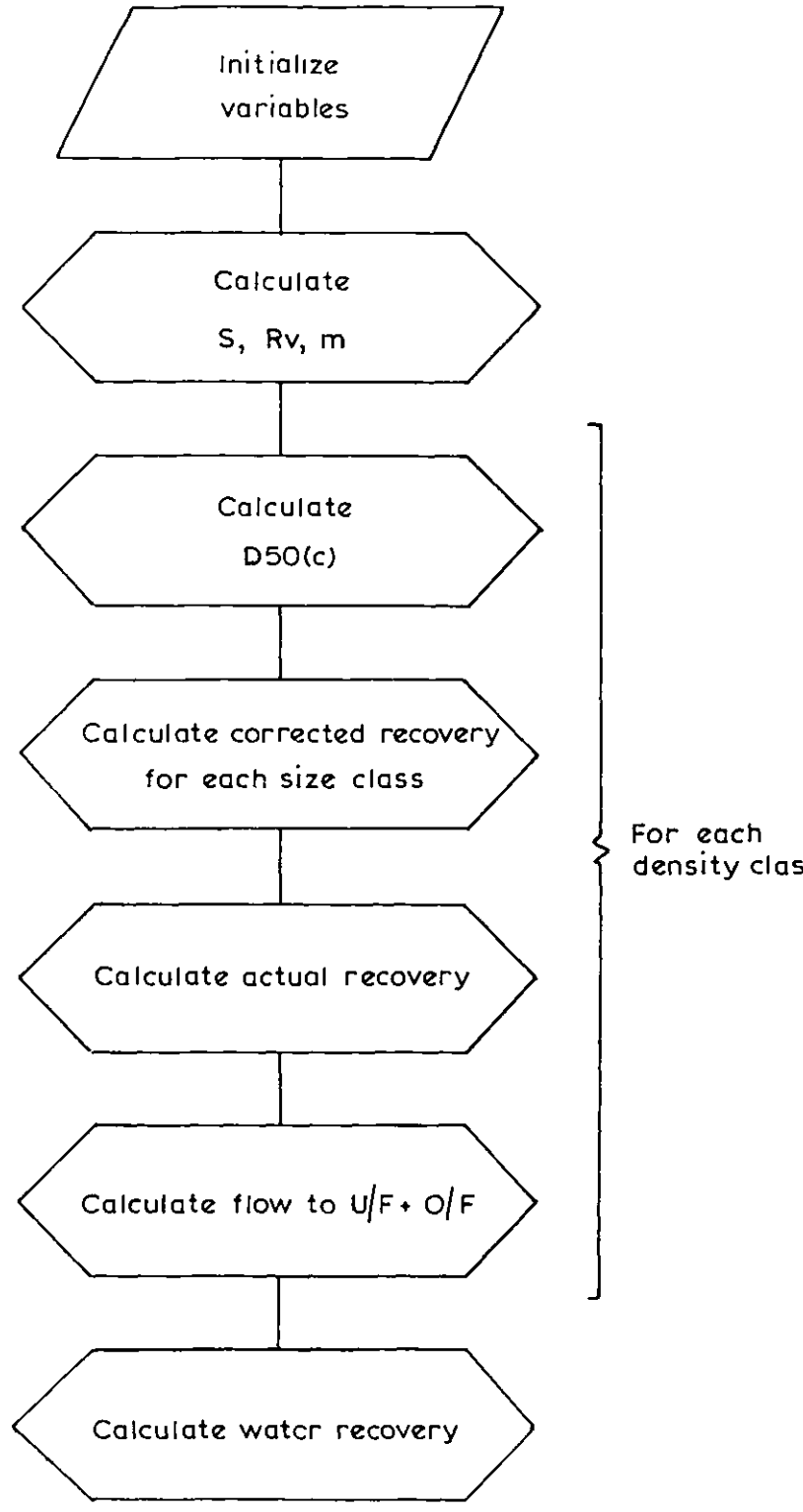


Figure 5.8 HYCYC model subroutine calculation procedure

number of hydrocyclones will be sufficient for the inlet flowrate also restricts the application and flexibility of the module, but in development it was felt that the advantages, in terms of simplifying model input and calculation procedure, justified these restrictions.

5.1.7 Flotation Modelling

A unit operation block, FLOT, was implemented on ASPEN PLUS to model the operation of a bank of flotation cells. The calculation procedure uses a first-order rate reaction analogy and is largely based on Sutherland's model (237, 239, 254). This formulation is likely to be of most use in design stages as it does not require many more parameters outside of the flotation rate constants. It is assumed that values for the rate coefficients are determined from testwork accurately reflecting the operating conditions (e.g. reagent regime, pulp density, machine characteristics) that will be found in the operating plant.

The model is based on the following assumptions:-

First-order kinetics

The feed material is assumed to consist of a number of fractions based on mineral composition and particle size. In each flotation cell the mass flowrate of the fraction passing to the concentrate is directly proportional to the mass of that component in the feed.

$$C_i = K_i . F_i$$

Perfect mixing

The pulp in the cell is assumed to be perfectly mixed.

Water flow

The split of water between tailings and concentrate is fixed by the definition of tailings pulp density.

Limit to rate of production of concentrate

It is necessary to allow for a maximum concentrate flow rate from a flotation cell and for a saturation of the bubble surface with solids at high concentrate flow rate. If cell concentrate flow rate (CTOT) exceeds a specified limit (AMAX) for this size of cell, operated in this manner, then the maximum solids concentrate flow rate (FMAX) is fixed and the flotation rates suppressed accordingly. FMAX is set by an arbitrary function (237) of AMAX to ensure a continuous system:-

$$FMAX = 2.AMAX \quad 1 - \frac{AMAX}{2.CTOT}$$

Whilst totally empirical, this function has been found to be valid in a number of applications and, rather than being a mechanistic description of the process, is intended to allow the user flexibility in scaling up testwork results.

A typical set of FLOT input data is shown in Fig. 5.9(a); the only equipment parameters which need to be defined are the number of cells (NO-OF-CELLS), the pulp containing volume of each cell (CELL-VOLUME), the cell tailings pulp density (TDEN) and the maximum cell concentrate flow (AMAX). It is also necessary to define the flotation rate constants for each particle size - mineral content fraction using the secondary keyword KVALUE. Typical block Input Data and Results report sections are shown in Fig. 5.9(b) and the calculation procedure is shown in Fig. 5.10.

5.1.8 Solid - liquid Separation Module

As was discussed in Section 2.3 there would appear to be two levels of detail which might be included in a model of solid-liquid separation operations. The more detailed approach uses theoretical and empirical relationships, which require considerable experimental parameter determination and is not well suited to use in a process feasibility study but is more appropriate for analysing operating plants; the model included in the commercial ASPEN PLUS system to simulate a rotary drum vacuum filter is of this type. As it was felt that this was too detailed for initial design purposes and as there was no module for simulating sedimentation processes, it was intended to model solid-liquid separation processes using the simple separation and mixing modules incorporated in ASPEN PLUS. For the use of such models, it would be necessary to have a knowledge of the separation and washing characteristics of the devices modelled and to assume that sufficient machine capacity was available to achieve that separation; this knowledge, and estimates of the necessary machine size, could be obtained from relatively simple laboratory testwork and accumulated industrial experience.

However a unit operation module was recently installed on a test version of ASPEN PLUS to simulate solids washing; whilst initially developed to simulate the settling tanks used in the Bayer process and based on the operating experience of alumina producing companies (197,

```
BLOCK B2 FLOT
  PARAM  NO-OF-CELLS=6    &
         CELL-VOLUME=20.0 &
         TDEN=2430.0      &
         AMAX=10.0
  KVALUE LEND=1 &
         SSID=NCPSMFD &
         KVAL=0.0072/0.0048/0.0022/0.0006/0.0003/
              0.0089/0.0064/0.0040/0.0009/0.0005/
              0.0110/0.0079/0.0058/0.0012/0.0009/
              0.0130/0.0115/0.0078/0.0018/0.0016/
              0.0140/0.0123/0.0097/0.0030/0.0023
```

Figure 5.9(a) FLOT Input language

```
FLOTATION    (FLOT   ):  B2
INLET = S2                OUTLET = S3          S4
  PROPERTY OPTION SET SYSOPO

                    ***  MASS AND ENERGY BALANCE  ***
                                  IN                OUT          RELATIVE DIFF.

   CONVENTIONAL COMPONENTS (KMOL/SEC)
      H2O                      11.0018           11.0018        -0.201825E-16
   SUBTOTAL(KMOL/SEC)          11.0018           11.0018        -0.201825E-16
          (KG/SEC   )          198.198           198.198         0.000000E+00
   NON-CONVENTIONAL COMPONENTS (KG/SEC   )
      METAL                    60.1398           60.1398        -0.590743E-16
      GANGUE                   111.439           111.439         0.000000E+00
   SUBTOTAL(KG/SEC   )         171.579           171.579        -0.414119E-16
   TOTAL BALANCE
      MASS(KG/SEC   )          369.777           369.777        -0.192154E-16
      ENTHALPY(WATT     )    0.369777E+38      0.369777E+38     -0.159635E-16

                         ***  INPUT DATA  ***

  CELL VOLUME              20.000
  TAILINGS PULP DENSITY    2430.0
  MAX. CONCENTRATE FLOW    10.000

  NUMBER OF CELLS          6

                         ***  RESULTS  ***

  COMPONENT 1     RECOVERY          GRADE

    CELL

      1          0.1420D+00        0.4467D+00

      2          0.1305D+00        0.4104D+00

      3          0.1253D+00        0.3966D+00

      4          0.1190D+00        0.3796D+00

      5          0.1118D+00        0.3604D+00

      6          0.1034D+00        0.3387D+00


        TOTAL              AVERAGE
     RECOVERY 0.7320D+00   GRADE 0.3892D+00

                      TAILING GRADE 0.2756D+00
```

Figure 5.9(b) FLOT Output

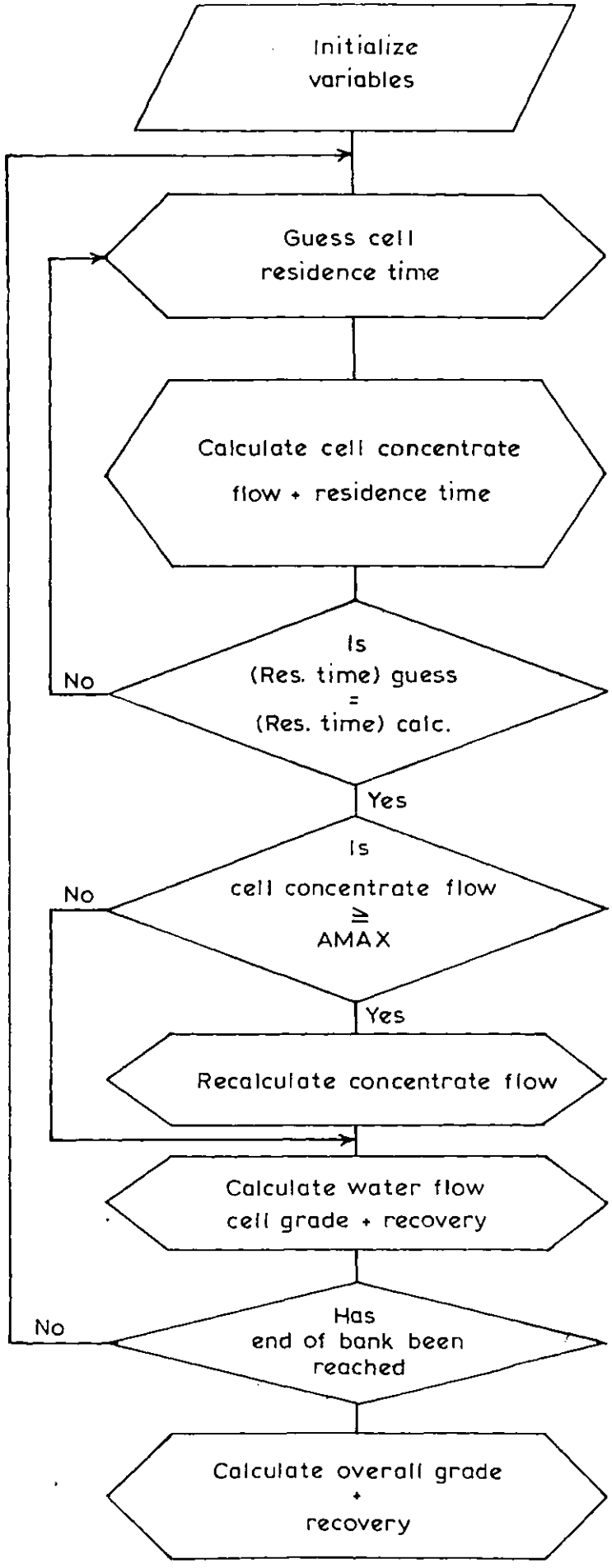


Figure 5.10 FLOT model subroutine calculation procedure

198), this has been implemented as a general purpose module whose application could be extended to include most solid-liquid separation operations in minerals extraction industries. The only necessary additions to the module would be equipment sizing and costing routines to allow it to be used to evaluate filtration and/or sedimentation operations and, possibly, its inclusion in a module to allow easier modelling of a complete installation rather than the individual separation devices. The calculation procedure used is based on the arrangement of mixers and separators shown in Fig. 2.10; thus it is possible to allow for both imperfect mixing and imperfect separation.

5.1.9 Agitation Leaching Module

The unit operation module LEACH was implemented on the ASPEN PLUS system to simulate the operation of a cascade of agitated leaching vessels arranged in a co-current fashion. It was concluded from a literature survey of the available models of such processes (see Section 2.5) that the phenomenological models, based largely on the shrinking core concept, were inappropriate for initial design purposes; the reasons leading to this conclusion included the need for excessive amounts of testwork to validate the use of a particular model and determine its parameters, the need to allow for more than one metal producing reaction and/or mineral, the need to allow for important side-reactions such as the lixiviant consuming gangue dissolution and the lack of experience in the reliable use of such models.

The module LEACH is therefore based on a completely empirical form and uses two different relationships to describe the consumption of lixiviant and extraction of metal value; the shape of these relationships is derived from the development of statistically based experimentation designs for the analysis of proposed agitation leach processes (309) and the mathematical forms used described in Table 5.2. Presently only one metal extraction is described, all other reactions being ignored in the mass balance calculations; thus the mass balance is calculated on the assumption that the only change to the solid phase is the removal of metallic ions to solution. Whilst it is realized that this assumption is invalid for most mineral-lixiviant systems it was felt during the development of the module that it would generally have little effect on the design and economics of the leach tanks and succeeding processing stages, unless the consequent increase in complexity of the leach model

```
BLOCK B1 LEACH
  PARAMETERS  NO-OF-TANKS=6     &
              TANK-VOLUME=600.0 &
              FRACTIVE-VOL=0.95 &
              BYPASS-FRAC=0.05  &
              MAKEUP-ACIDC=1000.0
  COEFF-A     0.0/0.0/0.0/0.0/0.0
  COEFF-B     -2.82/2.24/66.1/0.0/-247.0/0.0/1.66/0.0/-5.43/0.0/0.0/
              -156.0/0.0/0.0/0.0/0.0/0.0/-0.0184/-22.0/0.0/807.0/0.0
  COEFF-C     0.0/0.0/0.0
  COEFF-D     5.92/0.629/-17.1/0.0/-0.275/0.0/0.0/0.0012/11.4/0.0
  CONC-CONTROL  1 1 20.0  /
                2 1 30.0  /
                3 1 40.0  /
                4 1 40.0  /
                5 1 20.0  /
                6 0 0.0
```

Figure 5.12(a) LEACH Input language

*** RESULTS ***

TANK	EXTRACTION	RECOVERY	TOT.RECOVERY	TAILS GRADE
1	0.3403D+00	0.3403D+00	0.3403D+00	0.3675D-02
2	0.4216D+00	0.2781D+00	0.6184D+00	0.2125D-02
3	0.4354D+00	0.1661D+00	0.7846D+00	0.1200D-02
4	0.3655D+00	0.7873D-01	0.8633D+00	0.7613D-03
5	0.2422D+00	0.3310D-01	0.8964D+00	0.5769D-03
6	0.1509D+00	0.1563D-01	0.9121D+00	0.4899D-03

Figure 5.12(b) LEACH Output

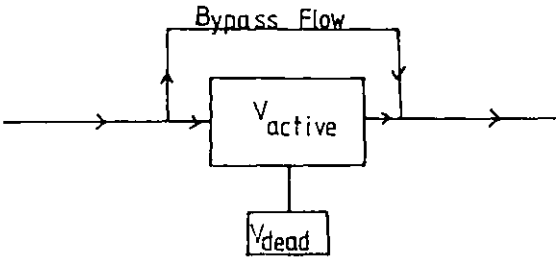


Figure 5.11 Cholette and Cloutier Mixing Model (458)

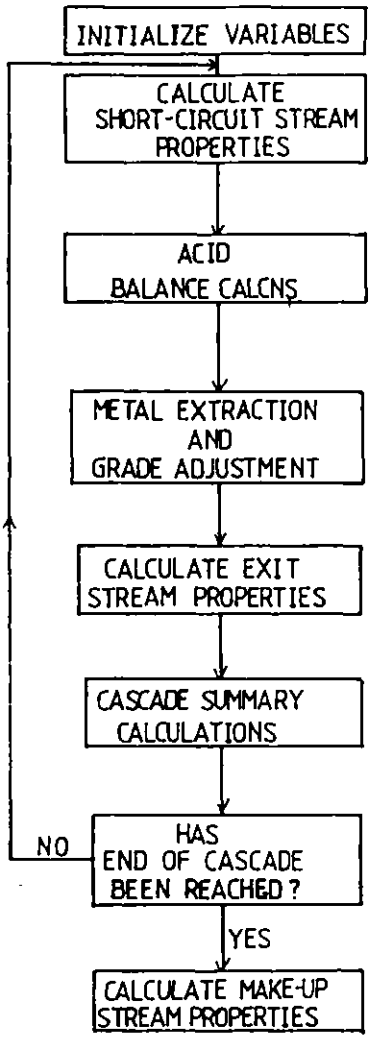


Figure 5.13 LEACH module model subroutine calculation procedure

Copper Extraction

X1 = CACID - A1 (CACID in g/l)
X2 = TIME - A2 (TIME in hours)
X3 = TEMP - A3 (TEMP in ^{o}C)
X4 = SIZ - A4 (SIZ in mm)
X5 = FSOL - A5 (FSOL in wt.fract.)

% Copper extraction = B0 + B1X1 + B3.X2 + B3.X3 + B4.X4 + B5.X5 +
B12.X1.X2 + B13.X1.X3 + B14.X1.X4 + B15.X1.X5
+ B23.X2.X3 + B24.X2.X4 + B25.X2.X5 +
B34.X4.X5 + B35.X3.X5 + B45.X4.X5 + B11.X1.X1
+ B22.X2.X2 + B33.X3.X3 + B44.X4.X4 + B55.X5.X5

Acid concentration

Y1 = CACIDI - C1 (CACIDI in g/l)
Y2 = TIME - C2 (TIME in hours)
Y3 = TEMP - C3 (TEMP in ^{o}C)

CACID = D0 + D1.Y1 + D2.Y2 + D3.Y3 + D12.Y.Y2 + D13.Y1.Y3
+ D23.Y2.Y3 + D11.Y1.Y1 + D22.Y2.Y2 + D33.Y3.Y3

Where CACID = solvent concentration in tank outlet stream
CACIDI = " " " " inlet "
TIME = mean residence time of pulp in tank
TEMP = mean pulp temperature
SIZ = particle size
FSOL = weight fraction of solids in pulp.

Table 5.2 : LEACH model equations.

was matched by more detailed modelling of subsequent stages; thus, for instance, if the dissolution of iron salts from the ore was modelled, the results of such modelling would need to be used to determine its effect on the solvent extraction and electrowinning stages, detail which was unlikely to be included in the modules implemented for these unit operations. Also only five major system variables are included in the empirical relationships; these were chosen based on experience (309) and are particle size, reagent concentration, temperature, pulp density and reaction time. If it was found desirable to include other variables such as pressure, it would be a relatively simple matter to make the necessary changes to the module coding and system definition file table.

Each agitated tank is modelled individually using the above relationships and the residence time distribution model proposed by Cholatte and Cloutier (458); the choice of this latter model is also somewhat arbitrary, but it was decided that it was suitable due to its simplicity and flexibility; there has been very little published work on the use of this or other models in leach tank design. As can be seen from Fig. 5.11, which shows the model schematically, the reaction is assumed to take place in a perfectly mixed tank whose volume is not completely used due to sanding out and which is bypassed by a fraction of material which is assumed to completely short circuit the tank. The addition of lixiviant to the cascade may take place in any of the individual tanks and is assumed to be controlled so that the concentration in that tank is fixed. From these concentrations and the calculation of reagent consumption, a mass balance around the cascade is determined, including the calculation of the total reagent make-up required.

Due to the nature of the second order empirical relationships used, the input of the extraction model coefficients is somewhat laborious; each of the coefficients shown in Table 5.2 must be input in order using the keywords COEFF-A, COEFF-B, COEFF-C and COEFF-D. However, as can be seen from Fig. 5.12(a) which shows a typical block input, the remainder of the parameter definition is somewhat less tedious; the residence time distribution is defined using the keywords NO-OF-TANKS (number of tanks in series), TANK-VOLUME, FRACTION-VOL (fraction of tank volume actively mixed) and BYPASS-FRAC (fraction of input flow effectively bypassing the tank. The addition of reagent is defined using the keyword MAKEUP-ACIDC to set the concentration of the reagent stream and the

of modules; however to retain some flexibility, the module should be restricted to simulating only a bank, rather than a complete solvent extraction plant of fixed configuration.

5.1.11 Copper Electrowinning Plant Simulation Module

As was discussed in Section 2.7, the modelling of each cell in a copper electrowinning tankhouse is generally unnecessary for initial design purposes and it is feasible to model the complete plant using simple relationships; although necessitating the treatment of the tankhouse as a "black box" with no consideration of the effect of internal cell arrangements, this approach has the advantages of complete simplicity and robustness whilst still providing reliable predictions of tankhouse performance and requirements. The major restriction of the approach is that it assumes that operating current efficiency and current density are known or can be estimated; the alternative to this assumption requires a considerably more detailed model of the electrodes and solution flows, and pilot scale testwork to determine model parameters.

Essentially all that is needed to simulate a tankhouse in this manner is a simple reactor model defining the rate of copper precipitation and acid production, these being determined from the available electrode area using Faraday's law. However in operating plants, the ambient conditions are controlled so as to ensure optimum performance; these conditions include the inlet and outlet copper and acid concentrations. Also the rate of copper production is dependent on the rate at which aqueous copper is added to the system in which the tankhouse operates; thus copper production is controlled by the efficiency of the leach and solvent extraction stages.

In the development of an ASPEN PLUS module to simulate the performance of a tankhouse, it was therefore decided to take into account the fact that for reliable prediction of current efficiency and density, the module user needed to be aware beforehand of the approximate ambient conditions. It was decided to allow the user to control the outlet stream flowrate and acid and copper concentrations, as well as defining the current efficiency and density; thus the calculation of the mass balance becomes a simple subtraction once the inlet stream properties are determined. From this mass balance it is then possible to calculate the necessary tankhouse size and power consumption for the amount of copper precipitated. If the opposite approach had been taken and the

tankhouse size was defined by the user, then the ambient conditions would have been more uncertain and an unstable simulation may have occurred if the tankhouse parameters were inappropriately defined. The relationships used to determine the tankhouse size and power consumption are those described in Table 2.3; in order for the cell voltage to be calculated it is necessary for the user to also define the average solution temperature and the electrode spacing.

5.1.12 Energy Balance Calculations

It is important to note that all of the unit operation modules implemented are only capable of calculating the mass balance of the operation; it was decided to include this simplification in the development of the models due to the extra complexity that energy balance modelling would have involved and because such calculations are unnecessary in many minerals extraction processes. A broad generalization would be that process energy balances are only important in a number of hydrometallurgical processes and are unimportant in nearly all physical and flotation separation processes. Therefore, the lack of an energy balance calculation capability is not a major deficiency, though it is one which needs to be remedied when more extensive simulation of hydrometallurgical processes is carried out.

5.2 Cost Estimation and Economic Evaluation

Due to the restricted time available in which to use the ASPEN PLUS system, it would have been impracticable to implement more than one unit operation cost module. It was therefore decided to develop a separate system with which to illustrate the cost and economic aspects of the methodology; however it should be noted that this should only be considered as a temporary measure and that use of the ASPEN PLUS CES subsystem is recommended for the future, due to its ability to interface with the process simulation routines.

The calculation procedures for evaluating the economics of a project are fairly standard once the major capital costs, operating costs and project revenues are known, and computer software is commonly available to carry out tasks such as determining overall cash flow net present values and internal rates of return, and carrying out sensitivity analyses. Whilst programs have also been written to estimate the capital and operating costs of mineral processing plants (24, 25), these are not as commonly available and are not integrated with economic analysis routines. Therefore routines were written to calculate basic plant costs and revenues and to input this data into an existing package for the economic analysis of mineral projects.

The program used as the core of this new system was the Mineral Economics Evaluation (MECON) package written by Dr. R. Spencer at the Royal School of Mines; the reasons for using this package included: the availability of advice on its use from the program's developer: its use of a flexible format data file as input, thus allowing it to be interfaced easily with the new routines: its proven and tested capabilities in analysing cash flows. Whilst MECON was devised as a flexible package able to adapt to any type of project by use of a separate data file containing the cost data and relationships, it was found that the data file was inadequate to handle a large variety of flowsheets costed using factorial techniques; it was therefore necessary to write a separate subprogram, PLANT, which reads data describing the plant, estimates basic plant costs and inputs these into the MECON data file for further processing.

The arrangement and interfacing of the various parts of the MECON/PLANT system are shown figuratively in Fig. 5.14; the documentation for the PLANT input data file and unit operation cost blocks is given in Appendix 2 of this thesis and the original MECON system has been

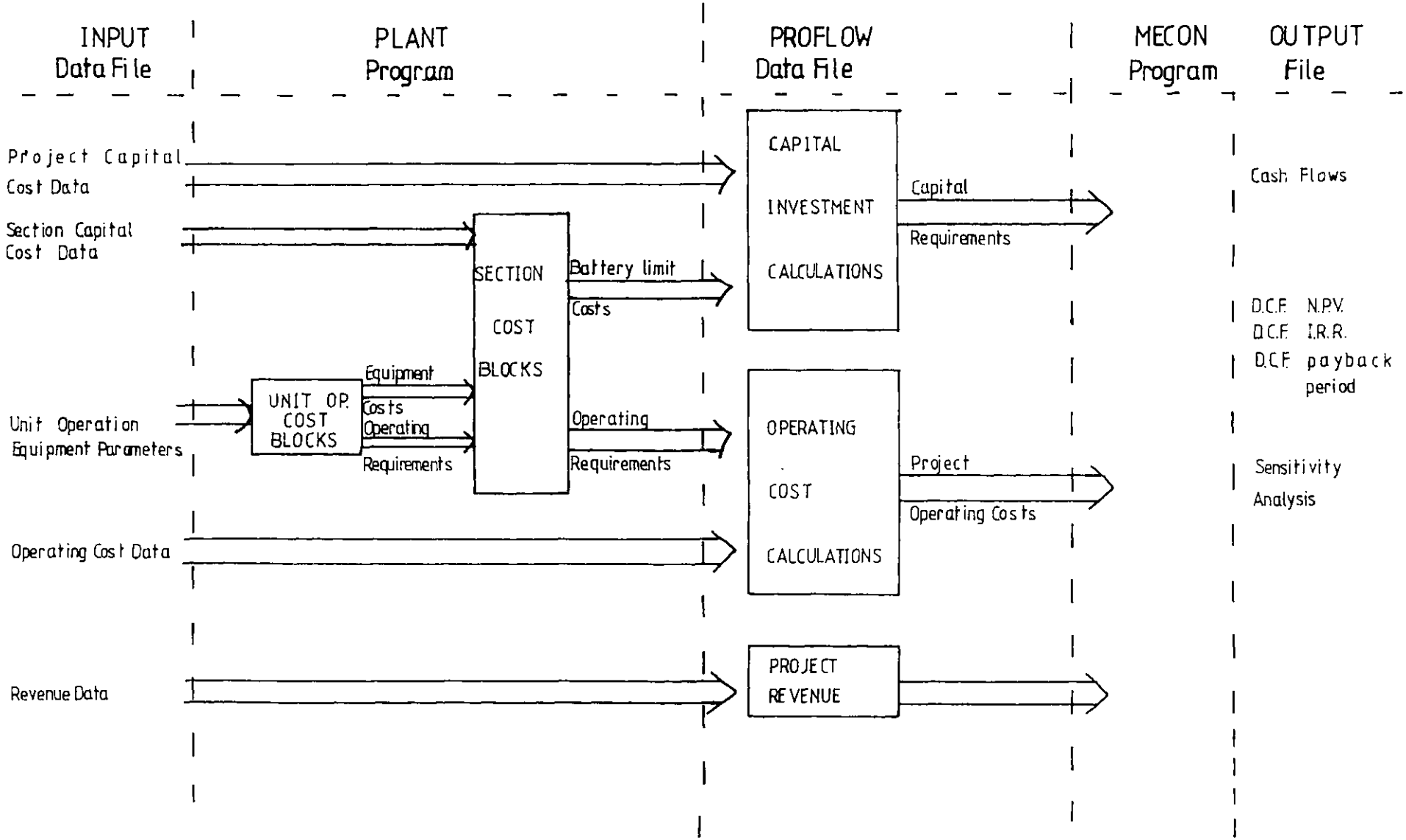


Figure 5.14 Shematic diagram of PLANT/MECON computer software system

documented by Cockerill (459). As can be seen from the diagram the input data for the system comprises of two parts, that data which refers to the project as a whole and is transferred directly into the PROFLOW data file, and the sectional data which is processed by the PLANT program before being transferred. Examples of the cash flow analysis and sensitivity analysis outputs are shown in Fig. 5.15.

Total plant capital investment requirements are estimated using Guthrie's sectional factorial technique (460); the actual calculation procedure is the same as used in the ASPEN PLUS system. Using this method a process plant is split into sections, within which similar unit operations are grouped together. To date, six plant sections have been included in the system, although the program structure can easily accept the introduction of new sections such as gravity separation; the six sections are crushing, grinding, flotation, solid-liquid separation, leaching and solvent extraction - electrowinning. For each section, "battery limit" installation material and labour requirements are estimated based on purchased equipment costs; these requirements include everything that is needed to construct the plant processing facilities. The cost of those extra facilities and requirements needed to support the process units are estimated during the projecting phase: following this phase, which determines the site requirements, the total capital investment needed is estimated, including the need for indirect costs, contingencies, working capital and start-up costs. The PLANT program calculates the section battery limit material costs and labour hours and the PROFLOW file is used by MECON to estimate the total capital investment.

To calculate the battery limit requirements of a process section it is first necessary to calculate the total purchase cost of the process equipment. This is carried out using unit operation cost blocks, the equipment listings of which are described in Appendix 2; the relationships used in these blocks are largely based on the cost information recently published by the CIMM (409). The program user needs only to input certain basic equipment parameters, the format for these also being described in Appendix 2. If the ASPEN PLUS system was being used the cost blocks would be of the same form, though the parameter specifications could alternatively be determined from the unit operation simulation blocks. Once the total section equipment cost has been determined, it is multiplied by a number of factors, as in Fig. 16 (a), to determine

MECON (MINERAL EVALUATION SYSTEM VERSION 2.1 - 83/2/24) DATE 83/04/25, PAGE 10

PROCESS PLANT FINANCIAL VALUATION MODEL

SUBROUTINE PLANT OUTPUT FILE ****NOTE .1000E+04 = 1000.0****

	YEAR 82	YEAR 83	YEAR 84	YEAR 85	YEAR 86	YEAR 87	YEAR 88	YEAR 89
-C- DATASET5	.1000E-01	.8700E+01	.1550E+01	.5700E+00	0	.3600E-01	.1130E+02	.4600E+01
-C- DATASET7	.1500E+02	.3000E+01	.2781E+07	.2659E+06	.1153E+05	.3951E+03	0	0
-C- DATASET9	.2100E+02	.1000E+01	.2793E+07	.1955E+06	.4320E+04	.3794E+06	0	0
-C- CRUSHING MAT COST	.8781E+07	0	0	0	0	0	0	0
-C- CRUSHING LAB HRS	.2249E+04	0	0	0	0	0	0	0
-C- GRINDING MAT COSTS	.1314E+08	0	0	0	0	0	0	0
-C- GRINDING LAB HRS	.3685E+06	0	0	0	0	0	0	0
-C- LEACH MAT COST	.6793E+07	0	0	0	0	0	0	0
-C- LEACH LAB HRS	.1566E+05	0	0	0	0	0	0	0
-C- SX-EW MAT COST	.2186E+08	0	0	0	0	0	0	0
-C- SX-EW LAB HRS	.2757E+05	0	0	0	0	0	0	0
-C-BATTERY LIMIT MAT COST	.5057E+08	0	0	0	0	0	0	0
-C-BATTERY LIMIT LAB HRS	.7797E+04	0	0	0	0	0	0	0
-C-TOTAL PLANT MAT COST	.5762E+08	0	0	0	0	0	0	0
-C-TOTAL PLANT LAB HRS	.9587E+06	0	0	0	0	0	0	0
-C-DIRECT MATERIAL COST	.5935E+08	0	0	0	0	0	0	0
-C-DIRECT LABOUR COST	.1680E+08	0	0	0	0	0	0	0
-C-FIELDCOSTS	.9623E+08	0	0	0	0	0	0	0
-C-TOTAL COSTS	.1087E+09	0	0	0	0	0	0	0
*C*TOTAL FIX CAP INV	.1434E+09	0	0	0	0	0	0	0
-C-ORE7COSTS	0	.1322E+08	.1322E+08	.1322E+08	.1322E+08	.1322E+08	.1322E+08	.1322E+08
-C-CONSUMMABLES COST 1	0	.2909E+06	.2909E+06	.2909E+06	.2909E+06	.2909E+06	.2909E+06	.2909E+06
-C-OPERATING LABOUR COST	0	.1282E+07	.1282E+07	.1282E+07	.1282E+07	.1282E+07	.1282E+07	.1282E+07
-C-POWER COST	0	.4000E+07	.4000E+07	.4000E+07	.4000E+07	.4000E+07	.4000E+07	.4000E+07
-C-CONSUMMABLES COSTS	0	.1811E+08	.1811E+08	.1811E+08	.1811E+08	.1811E+08	.1811E+08	.1811E+08
-C-COST OF LABOUR	0	.6232E+07	.6232E+07	.6232E+07	.6232E+07	.6232E+07	.6232E+07	.6232E+07
-C-OTHER OP COSTS	0	.1546E+08	.1546E+08	.1546E+08	.1546E+08	.1546E+08	.1546E+08	.1546E+08
-C-WORKING CAPITAL	.1460E+08	0	0	0	0	0	0	0
*C*REVENUE	0	.1305E+09	.1435E+09	.1578E+09	.1736E+09	.1910E+09	.2101E+09	.2311E+09
TOTAL ANNUAL SALES REVENUE	0	.1305E+09	.1435E+09	.1578E+09	.1736E+09	.1910E+09	.2101E+09	.2311E+09
***ORE COST	0	.1600E+08	.1760E+08	.1936E+08	.2129E+08	.2342E+08	.2577E+08	.2834E+08
***CONSUMMABLES COST	0	.2192E+08	.2411E+08	.2652E+08	.2917E+08	.3209E+08	.3530E+08	.3883E+08
***COST OF LABOUR	0	.7541E+07	.8295E+07	.9124E+07	.1004E+08	.1104E+08	.1214E+08	.1336E+08
***POWER COST	0	.4840E+07	.5324E+07	.5856E+07	.6442E+07	.7086E+07	.7795E+07	.8574E+07
***OTHER OPERATING COSTS	0	.1871E+08	.2058E+08	.2264E+08	.2490E+08	.2739E+08	.3013E+08	.3314E+08
TOTAL ANNUAL OPERATING COSTS	0	.6900E+08	.7590E+08	.8349E+08	.9184E+09	.1010E+09	.1111E+09	.1222E+09
***TOTAL CAPITAL EXPENDITURE	.1578E+09	0	0	0	0	0	0	0
***WORKING CAPITAL	.1506E+08	0	0	0	0	0	0	0
TOTAL ANNUAL CAPITAL COSTS	.1738E+09	0	0	0	0	0	0	0
UNADJUSTED CAPITAL ALLOWANCES	0	.1738E+09	0	0	0	0	0	0
ANNUAL TAXATION ALLOWANCES	0	.6145E+08	.6760E+08	.4477E+08	0	0	0	0
ANNUAL TAXATION AT 52.0 P.C.	0	0	0	.1353E+08	.4425E+08	.4876E+08	.5144E+08	.5661E+08
ANNUAL CASH FLOW	-.1738E+09	.6145E+08	.6760E+08	.7436E+08	.6641E+08	.4744E+08	.5217E+08	.5740E+08
CUMMULATIVE CASH FLOW	-.1738E+09	-.1124E+09	-.4477E+08	.2958E+08	.9599E+08	.1434E+09	.1956E+09	.2530E+09
PRESENT VALUES AT SELECTED DISCOUNT RATES								
DISCOUNT RATE 10.0 PERCENT	-.1580E+09	.5079E+08	.5079E+08	.5079E+08	.4123E+08	.2678E+08	.2678E+08	.2678E+08
DISCOUNT RATE 15.0 PERCENT	-.1511E+09	.4447E+08	.4445E+08	.4251E+08	.3302E+08	.2051E+08	.1942E+08	.1857E+08
DISCOUNT RATE 21.0 PERCENT	-.1437E+09	.4197E+08	.3816E+08	.3449E+08	.2560E+08	.1516E+08	.1374E+08	.1249E+08
DISCOUNT RATE 26.5 PERCENT	-.1374E+09	.3840E+08	.3339E+08	.2904E+08	.2050E+08	.1154E+08	.1007E+08	.8754E+07

INTERNAL RATE OF RETURN = 36.88 PERCENT

PAYBACK PERIOD - FROM FIRST REVENUE YEAR

UNDISCOUNTED VALUE	2.60 YEARS
DISCOUNT RATE 10.0 PERCENT	3.14 YEARS
DISCOUNT RATE 15.0 PERCENT	3.54 YEARS
DISCOUNT RATE 21.0 PERCENT	4.21 YEARS
DISCOUNT RATE 26.5 PERCENT	5.45 YEARS

Figure 5.15(b) Typical PLANT/MECON Cash Flow Output

MECON (MINERAL EVALUATION SYSTEM VERSION 2.1 - 83/2/24) DATE 83/04/29, PAGE 14

PROCESS PLANT FINANCIAL VALUATION MODEL

SUBROUTINE PLANT OUTPUT FILE ****NOTE .1000E+04 = 1000.0****

SENSITIVITY ANALYSIS ON PARAMETER - REVENUE £

FLEXED BY -	-50. P.C.	-30. P.C.	-10. P.C.	10. P.C.	30. P.C.	50. P.C.
ECONOMIC VIABILITY SENSITIVITY						
PAYBACK YRS.	++++++	6.04	3.21	2.19	1.66	1.34
	(PC)	(132PC)	(23PC)	(-16PC)	(-36PC)	(-49PC)
RETURN PERCENT	------	14.16	29.93	43.69	56.90	70.13
	(PC)	(-62PC)	(-19PC)	(18PC)	(54PC)	(90PC)
NPV AT 10.0 PC	-.2048E+09	.4051E+08	.2191E+09	.3927E+09	.5648E+09	.7362E+09
	(-157PC)	(-97PC)	(-28PC)	(28PC)	(84PC)	(141PC)
NPV AT 15.0 PC	-.1831E+09	-.6363E+07	.1239E+09	.2486E+09	.3714E+09	.4935E+09
	(-198PC)	(-103PC)	(-34PC)	(33PC)	(99PC)	(165PC)
NPV AT 21.0 PC	-.1652E+09	-.3940E+08	.5521E+08	.1443E+09	.2314E+09	.3178E+09
	(-265PC)	(-139PC)	(-45PC)	(44PC)	(131PC)	(219PC)
NPV AT 26.5 PC	-.1533E+09	-.5713E+08	.1676E+08	.8552E+08	.1522E+09	.2182E+09
	(-398PC)	(-211PC)	(-67PC)	(66PC)	(196PC)	(325PC)

Figure 5.15(b) Typical PLANT/MECON Sensitivity Analysis Output

a) $INMAT = \left(\sum_{i=1}^{i=8} mFi \sum UC_j \right) + \sum UC_j$

$INLAB = \sum_{i=1}^{i=9} lFi \sum UC_j$

Installed material cost
Installed labour hours
where UC = unit operation purchased cost

i	Material factor (mFi)	Labour factor (lFi)
1	Piping	Piping
2	Instrumentation	Instrumentation
3	Foundations	Foundations
4	Structure	Structure
5	Insulation	Insulation
6	Electrical	Electrical
7	Pointing	Pointing
8	Miscellaneous	Miscellaneous
9	-	Installation

b) $BLMAT = \left(\sum_{i=10}^{i=14} (mFi)\ INMAT \right) + INMAT \times \frac{CI_T}{CI_B}$

$BLLAB = \sum_{i=10}^{i=12} (lFi).INMAT + INLAB$

Battery limit material cost
Battery limit labour hours
where CI_T = Present cost index
CI_B = Base cost index

i	Material factor (mFi)	Labour factor (lFi)
10	Building	Building
11	Testing	Testing
12	Add. unaccounted	Add. unaccounted
13	Capital Spares	-
14	Unlisted equipment	-

c) $SBMAT = mF15.BLMAT$ — Service building material costs

$SBLAB = lF15.SBMAT$ — Service building labour hours

$SSMAT = mF16.(BLMAT + SBMAT)$ — Service system + utility distributi costs

SSLAB = 1F16.SSMAT — Service system + utility distribution hours

ADMAT = mF17.BLMAT — Additional direct material cost

ADLAB = 1F17.ADMAT — Additional direct labour hours

TDMAT = BLMAT + SBMAT + SSMAT + ADMATT — Total direct material costs

TDLAB = BLLAB + SBLAB + SSLAB + ADLAB — Total direct labour hours

d) SDMAT = mF18.TDMAT — Site development material cost

SDLAB = 1F18.SDMAT — Site development labour hours

FREIGHT = pF_1.TDMAT — Freight + delivery costs

ASMCOST = TDMAT + SDMAT + FREIGHT — At-site direct material cost

ASLCOST = LCOST.LPROD.(TDLAB + SDLAB) — At-site direct labour cost

where LCOST = Construction Labour Payrate
LPROD = Construction Labour Productivity

PDIRCOST = ASMCOST + ASLCOST — Project Direct Cost

e) FINMAN = pF_2 . ASLCOST — Field Indirect Material Cost

FINLAB = pF_3 . ASLCOST — Field Indirect Labour Cost

FLDCST = PDIRCOST + FINMAT + FINLAB — Field Construction Cost

f) CPROJMAN = pF_4 . FLDCST — Project management cost

CENGHOFF = pF_5 . FLDCST — Engineering and home office cost

CFEEPERM = pF_6 . FLDCST — Fees, permits and insurance cost

CADDCOST = pF_7 . FLDCST — Additional depreciable cost

DEPCOST = FLDCST + C PROJMAN + CENGHOFF + CFEEPERM + CADDCOST

DEPCOST = Total depreciable cost (excluding contingencies

g) PROCONT = pF_8 . DEPCOST — Process contingency

PRODEFF = pF_9 . DEPCOST — Project definition

TDCINV = DEPCOST + PROCONT + PRODEFF + CLAND — Total Depreciable Capital Investment

Figure 5.16 : Capital Investment Estimation Procedure

the total installed material cost and construction labour hours for the section process equipment; as can be seen from the diagram, these factors allow for the cost of items such as piping, electrical systems and instrumentation. If desired, the program user may over-ride any or all of these factors' default values and input the actual cost, labour hours or a different factor; the default values for all factors are given as part of the program documentation in Appendix 2. To determine the section battery limit requirements, the installed material cost is multiplied by a further set of factors as in Fig. 5.16 (b) to account for building costs etc.

The calculation of the total capital investment, based on the total plant battery limit costs and labour hours, is shown in Fig. 5.16 (c)-(g); this is a stepwise procedure relying again on the use of factors. Initially the project direct costs and field construction costs are estimated (Fig. 5.16 (c)-(e)),these being the costs involved with constructing the process plant and its ancillary buildings and systems and maintaining the construction labour force at the site. Following these steps, the project indirect costs and allowances are estimated (Fig. 5.16 (f)-(g)), these including the costs involved with planning and managing the project and contingency allowances to cover costs which may be incurred due to uncertainties and inaccuracies arising during project planning.

Calculation of the process operating costs is split into five sections; ore, labour, power, consummables and other operating costs. Whilst a minerals extraction plant is generally run in close association with one or more mines and the costs of ore processing are considered as an integral part of the overall project economics, it is useful when studying the process plant economics to have an estimate of the effective cost of the mined ore so that the effects of taxation on project cash flows may be understood. In this way decisions may be more reliably made on, for instance, the choice between a low cost, low recovery process, such as dump leaching, and a high cost, high recovery operation, such as flotation; similarly the balance between higher capital expenditure and lower operating costs may be judged more reliably. Ideally cost models of the project units, analagous to the process sections, would be included and would allow for mine capital and operating costs, downstream operations such as smelters and refineries and project

overheads such as power generation or transmission, roads and infrastructure; examples of similar relationships have been provided by O'Hara so as to provide approximate cost and cash flow figures. It should be noted however that some care and experience is required for suitable models to be implemented and in their use; also the amount of detail to be included is important in many cases, such as the siting of leach dumps and therefore the costs of ore haulage to the dumps. In the absence of suitable models and the likely inexperience of process engineers with respect to mine engineering and economics, an alternative method of including a suitable figure is to use an estimate of the overall cost per tonne of ore mined; whilst this is likely to err on the side of conservatism, due to ignoring capital tax allowances, this is perhaps better than allowing the potential of totally misleading mining cost estimates due to misuse of the models provided. Using such an estimate it is possible to compare relative process economics and to determine the most attractive process results, taking into consideration capital expenditure, taxation, etc., though it is impracticable, or at least inaccurate, to determine an absolute value for the profitability of a process; this is however an area where a great deal of further discussion, study and development work is needed so that the methodology for process evaluation may be efficiently incorporated into project evaluation.

The costs of power consumption and consummables, such as comminution media and reagents, are estimated from the sum of unit operation requirements and user inputted unit costs. In the case of consummable goods it is expected that the unit costs of the materials will take into consideration any extra freight, handling and storage charges that might be incurred and will thus reflect their true cost rather than market cost. Power costs are treated in a similar fashion although it is realised that this method is rather simplistic and could lead to inaccuracies; it would be advisable when developing the ASPEN PLUS costing system to include allowances for costs incurred in voltage transformers and rectifiers, peak loads and ancillary use of power in lighting, heating etc. It is felt however that using the maximum power rating of the installed motors, an approximate and probably conservative estimate of power consumption should be obtained.

The cost of labour is obtained using a factorial method, as is the estimation of other operating costs, such as maintenance labour and supplies; the calculation procedures are shown in Fig. 5.17 (a) and (b). As can be seen the overall labour costs, including supervision, administration and payroll burdens, are based on the number of process operators and their average hourly pay rate. The use of a factor to allow for supervisory staff eliminates the need for producing a detailed plant manning chart, whilst factors are almost the only realistic method of accounting for the overheads associated with maintaining a workforce; values for these factors should be relatively easy to determine from plant records, though it should be noted that the overhead costs are likely to be considerably higher for an isolated project with living facilities provided, than for a project where supporting infrastructure is already available. The calculation procedure for estimating other operating costs is largely based on the total fixed capital investment involved in building the plant and is a fairly commonly used method, so that factors should again be readily available.

All of the calculation steps involved in estimating the total operating costs are defined in the PROFLOW data file and carried out by MECON; the only relevant calculations made in the PLANT program are to estimate the consumption of media and reagents and to calculate the total installed motor power ratings. Similarly the estimation of project revenues and working capital requirements is handled by MECON, as are the calculation of tax paid, inflation and overall cash flows. These are then evaluated using discounted cash flow techniques, including net present value, internal rate of return and payback period.

Cost of Consumables = Sum of major plant media and reagent consumption cost.

Power Cost = (Sum of installed equipment maximum power ratings). (Unit power cost)

Cost of labour = OCF_1.(No. of operator man hours.pay rate + OCF_2.TFCINV)

Other Operating Costs = OCF_3.(total operator pay) + (OCF_4 + OCF_5).(Cost of labour)
+ TFCINV.(OCF_6 + OCF_7 + OCF_8 + OCF_9)

where TFCINV = total fixed plant capital investment and:-

Operating Cost Factor	Purpose
OCF_1	Labour supervision
OCF_2	Maintenance labour
OCF_3	Operating labour supplies
OCF_4	Administration
OCF_5	Payroll burden
OCF_6	Maintenance labour supply
OCF_7	Local taxes
OCF_8	Local insurance
OCF_9	Extra operating costs

Figure 5.17 Operating cost calculation procedure.

5.3 "Stand-alone" Process Simulation Programs

In a similar fashion to the restricted availability of the ASPEN PLUS system resulting in the development of the PLANT/MECON system to demonstrate the costing and economic aspects of the methodology, it was also impractical, though also unnecessary, to demonstrate the use of the implemented modules for mass balance calculations. As is described in Sections 5.1 and 6.4, the modules were used on the ASPEN PLUS system and the calculated results checked against hand-calculated model solutions to ensure correct installation; a number of the modules were also used in a restricted set of flowsheet simulations to test their robustness in such situations and to gain experience in the use of mineral processing models on the ASPEN PLUS system. However it was impractical to carry out extensive tests on the modules and so a number of "stand-alone" simulation programs were written, using the module model subroutines, to test the response of the models to changes in paramater values and to demonstrate their use in process simulation. Whilst using the ASPEN PLUS model subroutines, these programs had their own parameter definition, flowsheet solution, unit operation costing and report writing sections and so could be used independently of the ASPEN PLUS system. Each program had an array stream data structure of the type used in the model routines so that it was possible to dispose with the plex data structure, stream data handling subroutines and module interface routines; the major restriction in using these "stand-alone" programs was that a separate program needed to be developed for each flowsheet simulated.

The three main programs written modelled the flowsheets shown in Figures 5.18 -5.20 ; similar programs were also written to individually evaluate the crusher, screen, hydrocyclone, ball mill, flotation and leach models. The structure of each of the flowsheet programs is shown in Figures 5.18 -5.20 , the smaller programs having a similar structure except for not needing the flowsheet solution section. The subroutine BWEG used to accelerate the convergence of the model solutions was adapted from the appropriate ASPEN PLUS subroutines and uses the bounded Wegstein method. Due to the temporary nature of the programs, elaborate input and output routines were not developed with parameter definition taking place within the program and the output simply consisting of a listing of each of the stream constituents and overall capital and operating costs.

```
        CALL  INITIAL                          (Variable initialization)
        CALL  SCREEN2(1)
        CALL  SCREEN2(2)
        CALL  MIXER  (1)
        CALL  CRUSH2 (1)
        CALL  MIXER  (2)
 250    CALL  SCREEN2(3)
        CALL  SCREEN2(4)
        CALL  MIXER  (3)
        CALL  CRUSH2 (2)
        CALL  MIXER  (2)
        CALL  BWEG   (IFLAG)
        IF  (IFLAG.LT.1)   GO TO 250          (Test convergence)
        CALL  MIXER  (4)
        CALL COST                          (Capital and operating cost estimation)
        CALL WRIT                              (Report writing)
```

Figure 5.18 Crushing simulation program subroutine calling executive.

```
        CALL INITIAL
        CALL MIXER   (1)
250     CALL GMILL
        CALL MIXER   (2)
        CALL HYCYC2
        CALL MIXER   (1)
        CALL BWEG  (IFLAG)
        IF (IFLAG. LT.1 ) GO TO 250
        CALL COST
        CALL WRIT
```

Figure 5.19 Grinding simulation program subroutine calling executive.

```
        CALL INITIAL
        CALL MIXER
250     CALL FLOT   (1)
        CALL FLOT   (2)
        CALL FLOT   (3)
        CALL MIXER
        CALL BWEG  (IFLAG)
        IF (IFLAG.LT.1)  GO TO 250
        CALL COST
        CALL WRIT
```

Figure 5.20 Flotation simulation program subroutine calling executive.

CHAPTER 6 Case Studies Illustrating the Application and Implementation of the Methodology

As has been described in the previous chapter, the time available to use the simulation capabilities developed around the ASPEN PLUS system was severely restricted and so it was necessary to implement further programs to test and illustrate the applications, and limitations, of the techniques on which the proposed methodology is based. The objectives of this initial testwork were fourfold:-

i) to test the models' robustness in simulations

ii) to test the sensitivity, and reliability, of the models to model parameters and inherent assumptions

iii) to illustrate the type and variety of data produced by the methodology

iv) to gain some experience in the use of the proposed techniques, especially in terms of their limitations and any potential improvements which may be found to be suitable.

The major obstacle to carrying out relevant testwork is the almost complete lack of suitable published data, either for simulating the physical process or for evaluating the economics of a process. Whilst perhaps it is not surprising that there is no data for studying process economics based on process modelling, due to the lack of similar work, it is worthy of note that very few models described in the technical literature for process simulation are accompanied by comprehensive data supporting their use and application; although it could be argued that the use of the models is unjustified without such supporting evidence, it has been assumed that future testwork to verify the methodology will prove the validity of the models and/or that the models will be adjusted accordingly. For the purposes of this initial implementation and testing of the methodology it was therefore necessary to use approximate data in hypothetical case studies and to test the sensitivity of each system's responses to changes in the system input and the model variables.

In addition to the work carried out on the ASPEN PLUS system, which was of a subjective nature and concerned with the capabilities and limitations of that system in a minerals extraction application, 3 case studies were carried out using the programs developed separately. The first of these was concerned with the cost and economic aspects of a

base metal flotation operation rather than the processing, whilst the other two were more concerned with using process models to study performance and how this was related to costs; these latter dealt with another flotation operation and a hydrometallurgical extraction respectively

6.1 Economic Evaluation of a Copper-Molybdenum Flotation Plant

Using a detailed description, including a full equipment list, of the Sierrita copper-molybdenum concentrator (461), an input data file for use with the PLANT/MECON system was developed so that a cost and economic evaluation could be carried out. A summary of the conditions used as the baseline for the study is shown in table 6.1 and the resulting economic output is shown in table 6.2; all of the input data was extracted from the description of the Sierrita plant except for the mining cost per tonne, the project lifetime, plant operating schedule, product prices, media consumption and costs, and the capital and operating cost estimation factors.

The mining cost per tonne was estimated using O'Hara's relationships (412), as updated by Mular (409), and, whilst conservative, is typical of a high tonneage, open-pit operation for moderately difficult conditions; the nominal cost per tonne is made up of a capital cost element, amortized over 15 years at 10%, and an operating cost element. As discussed in the previous chapter, this method of allowing for the "delivered" cost of the ore is unsatisfactory if an absolute value for economic factors such as net present values is needed, but does allow more reasonable estimates of actual costs and cash flows for use in comparative studies. The values used for estimating the cost of media consumption are based on average industry-wide figures as published by the U.S.B.M. (462). The cost estimation factors used are the default values used in the PLANT/MECON system (see Chapter 5 and Appendix 2), whilst the product prices, considerably higher than April 1983 levels, were adjusted to provide a positive internal rate of return. General inflation was assued to be 10% per annum and a taxation schedule similar to the United Kingdom's was used, though, of course, a more accurate Canadian schedule could have been used.

The output information shown in table 6.2 is a summary of the information given in the present system's report; whilst only the total installed material cost and labour hours for each plant section is printed, it is feasible to output cost data for each block, an option which should be provided if the ASPEN PLUS cost estimation subsystem is used. Similarly only total plant power consumption and consummables costs are printed, though this is only dependent on presentation style and more detailed information may easily be provided. Hand calculated

Location: Montreal

M + S (MM) Index at start-up: 800

Project lifetime: 15 years Average Annual Inflation: 10%

Taxation: 52.0% annually : 100% Capital Allowances

Operating schedule: 49 weeks annually

Mining costs/tonne: $1.94 ($1.05 Capital cost element)
($0.89 Operating cost element)

Feed tonneage: 85,000 tpd (7 days per week)

Plant products:

Copper concentrate	:	1,735,223 kg/week at $1.9/kg
Moly concentrate	:	126,264 kg/week at $14.0/kg
Silver in concentrates	:	410 kg/week at $240.0/kg

Cost of consummables:

Crushing steel	$1.55/kg
Grinding steel	$0.57/kg
Flotation reagent	$0.1071/kg
Power	$0.03/kWh
Operating labour	$8.00/hr (basic pay)

Crushing Section

4 operators/shift : 18 shifts/week

2 primary gyratory crushers 2.26 m x 1.47 m : 373 kW motors
1 primary conveying system 4 km long : 1.37 m wide
8 coarse ore bins : 9200 tonnes capacity
4 scalping screens (double deck) 1.83 m x 3.66 m
4 secondary cone crushers 2.55 m : 260.75 kW motors
4 secondary screens (single deck) 1.83 m x 4.88 m
10 surge bins : 203 cu.m
10 tertiary cone crushers 2.1 m : 298.0 kW motors
10 tertiary screens (single deck) 1.83 m x 4.88 m
8 fine ore bins 9200 tonnes capacity
various connecting conveyor belts

Grinding Section

3 operators/shift : 21 shifts/week

16 primary ball mills 5.03 m x 5.79 m : 2235 kW motors
16 hydrocyclone clusters : 5 x 0.87 m diameter cyclones in each

Table 6.1 : Copper-Molybdenum Plant Base Conditions

(all costs in Canadian dollars)

(Continued)

2 regrind ball mills 3.35 m x 4.60 m : 596 kW motors
2 hydrocyclone clusters : 8 x 0.254 m diameter cyclones in each

Flotation Section

3 operators/shift : 21 shifts/week

48 rougher banks	:	10 x 2.83 m^3 cells each
14 scavenger banks	:	9 x 1.132 m^3 cells each
6 cleaner banks	:	6 x 1.132 m^3 cells each
4 moly rougher banks	:	12 x 2.83 m^3 cells each
10 moly cleaner banks	:	12 x 1.132 m^3 cells each
1 moly cleaner bank	:	4 x 1.132 m^3 cells

Solid-Liquid Separation Section

3 operators/shift : 21 shifts/week

5 rotary drum filters : 47.0 sq.m area each
5 thickeners : 106.8 m diameter
1 Copper concentrate handling facility
1 Moly concentrate handling facility
1 Tailings dam

Table 6.1 (Continued)

Capital Costs

Battery limit material cost	$108.7 m
Battery limit labour hours	$ 2.846 m
Total fixed capital investment	$363.7 m
Working capital	$ 29.5 m

Operating Costs (1st operating year)

Consummables cost	$ 26.79 m
Cost of labour	$ 15.59 m
Power cost	$ 20.02 m
Other plant operating costs	$ 44.63 m
	$107.03 m
Mining costs	$ 68.55 m
	$175.58 m

Revenue (1st operating year) - $306.1 m

Net Cash Flow (1st operating year) + $130.5 m (no tax payments until 4th operating year)

Internal Rate of Return $32.18% (NB: inflation at 10%: real rate = 20.2%

Net Present Value (21.0%) $172.8 m

Payback period (undiscounted) 3 years

Table 6.2 : Copper-Molybdenum Project : Base Economic Conditions

values for the base conditions are shown in Table 6.3 to provide this higher level of detail and include the operating costs for each section. The potential use of such a report is apparent when it is shown that the power consumption in the flotation section is three times greater than the crushing section, highlighting one of the inefficiencies of the older, smaller flotation machines. Four limitations of the present costing methods became apparent when used with the detailed information available; a common type of electrical supply was assumed, which is inappropriate for the variety of motors and other electrical appliances to be found around a minerals extraction plant: power consumption is based on the maximum rating of each motor, rather than typical operating ratings, and does not include power consumption other than by major process equipment: more detailed categorization of consummables is possible, such as grinding media and liners, and different flotation reagent types: allowance must be made in the input file for the extra shift operators off duty each day. Despite these possible limitations, the methods used do provide a reasonable estimate of the process costs from a minimal amount of input data; the possible limitations outlined could only be remedied by increasing the amount of input data necessary. Table 6.3 also indicates another limitation of the operating cost estimation method used, this being its dependency on the use of factors to estimate a large proportion (55.3%) of the total operating cost; Figure 6.1 illustrates the considerable sensitivity of the total operating costs to these factors. The values used in this evaluation are the default values of the PLANT/MECON system, although these are obtained from chemical engineering sources (460) and hence may be inappropriate.

Table 6.4 and Figure 6.2 illustrate how dependent the estimation of the total fixed capital investment also is on the correct application of cost estimation factors; in this study the total capital cost is 4.75 times larger than the delivered equipment cost upon which its estimation is based, although Figure 6.2 shows that the necessary investment is most sensitive to the equipment costs rather than to the cost estimation factors. In this instance the factors used were the default values for the cost estimation system, the installation cost factors being derived from mineral industry sources, though the project and ancillary cost factors are from chemical industry sources,

	$m	% of total
Crushing Power	1.49	1.38
Liners	1.96	1.83
Grinding Power	13.55	12.66
Media	16.79	15.69
Flotation Power	4.76	4.45
Reagents	8.05	7.52
S/L Sepn Power	0.22	0.21
Total Power	20.02	18.71
Total Consummables	26.8	25.04
Operating labour pay	0.99	0.92
Maintenance labour pay	12.00	11.21
Supervision costs	2.60	2.43
Operating supplies	1.87	1.75
Maintenance supplies	14.40	13.45
Administration cost	9.35	8.74
Payroll burden	6.24	5.83
Other operating costs	12.76	11.92
	107.03	100.0

Table 6.3 : Breakdown of Process Operating Costs in 1st Year

(Canadian dollars)

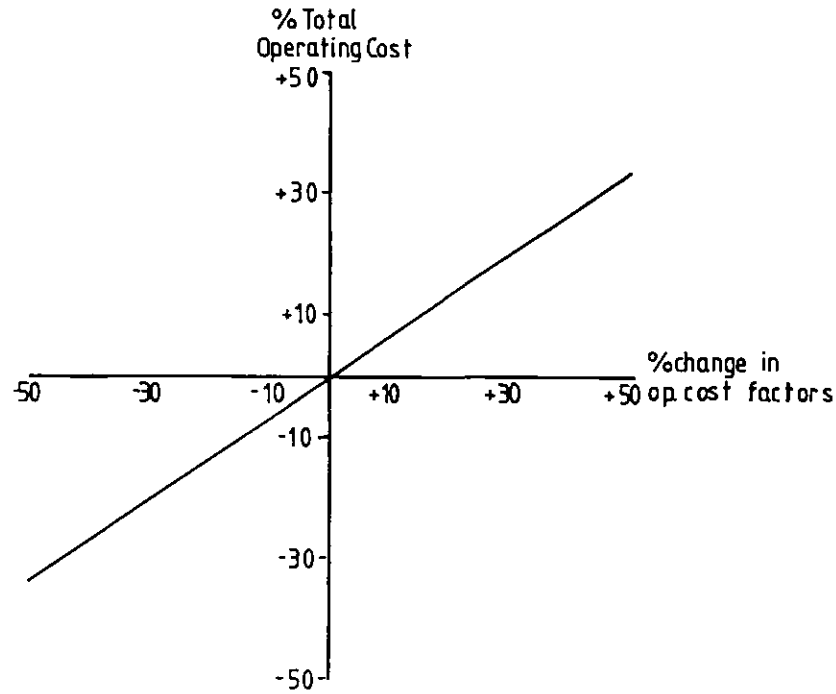


Figure 6.1 Sensitivity of Process Operating Costs to Factors

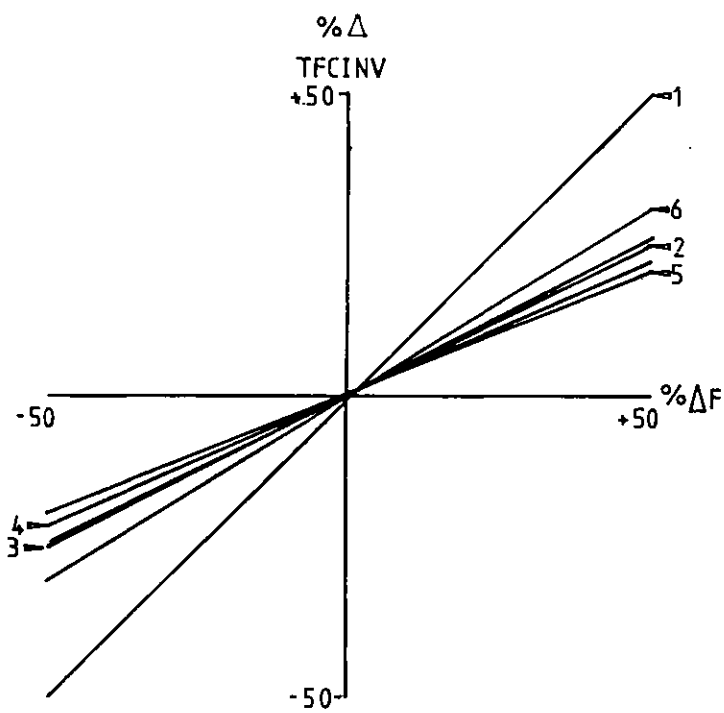


Figure 6.2 Sensitivity of Total Fixed Capital Investment to Changes in Cost estimation factors

Key F = 1 Delivered equipment cost
2 Installation cost factors
3 Buildings + services cost factors
4 Project cost factors
5 Construction labour payrate
6 Cost index

	$m	% of TFCInv
Crushing equipment	20.2	5.55
Crushing installation material	6.58	1.81
Crushing Installation labour	12.0	3.3
Grinding equipment	38.4	10.56
Grinding installation material	13.65	3.75
Grinding installation labour	25.6	7.04
Flotation equipment	11.4	3.13
Flotation installation material	7.72	2.12
Flotation installation labour	8.0	2.2
S/L Sepn. equipment	6.5	1.79
S/L Sepn. installation material	4.49	1.23
S/L Sepn. installation labour	4.6	1.26
Service buildings + systems material	15.2	4.18
Service buildings + systems labour	6.74	1.85
Freight + site development	3.72	1.02
Indirect field costs	34.2	9.4
Project management, engineering, home office, fees, permits etc.	56.5	15.53
Process contingency and definition, land and royalties	88.2	24.25
Total fixed capital investment	367.7	100.0

Table 6.4 : Breakdown of Capital Cost Estimate

though the project and ancillary cost factors are from chemical industry sources. These latter account for 56% of the estimated total fixed capital investment and their use should therefore be validated before they are used extensively, even though Figure 6.2 shows that the total figure is no more sensitive to these figures than the more reliable installation cost factors.

It can be seen from table 6.2 that the project economics at the base conditions appear to be extremely attractive, though it should be noted that the economic indicators must be treated with caution as the plant should be considered as an integral part of a larger mineral project. The manner in which mining costs are handled is an obvious example of this, especially as these are shown to have a significant effect on the project economics in figure 6.4; also, possible extra project costs associated with the local infrastructure have been ignored. Whilst these process economic evaluations are useful studies with respect to the plant design and operation, it would be useful at a later date to determine how the plant cash flows may be incorporated more reliably into the overall project evaluation, perhaps through the use of common output/input data files or even including more detailed mining and project cost blocks as an integral part of the ASPEN PLUS system.

The main value of an economic model of a process, other than as part of the project evaluation, is in studying the relative effects of important process and economic variables. Once the basic model has been set up on a computer, it is a fairly easy and quick task to change model variables and study their effects on the overall project; with a program such as PLANT/MECON the task is made even simpler by the automation of sensitivity studies on designated variables. Using these methods, it is a fairly easy task to identify the cost sensitive areas and it is these which most design effort ought to be concentrated on. The results of sensitivity studies carried out on this particular model are summarized in Figures 6.3 and 6.4. From Figure 6.4 it can be seen that the project's economic viability is most sensitive to the annual revenue and to the annual operating costs. If an estimated positive net present value is taken as the criterion for viability, then a 17% drop in process revenues could be sustained before this condition is reversed; such a drop may result from lower ore or

concentrate grades, metal recovery, product prices and/or a combination of these. Assuming that all other conditions remain constant the project would appear to require a copper price of $2.0/kg (91c/lb) (if smelter charges are $0.50/kgCu) and a molybdenum price of $11.4/kg.

A 29% decrease in total operating costs would also achieve breakeven economics, though it must be stressed that these are themselves highly sensitive to the apparent mining costs, as can be seen in Figure 6.4. It is also important to note that the second most important operating cost section, the miscellaneous costs, is, after mining costs, the least reliably calculated cost due to its dependence on cost factors in its calculation.

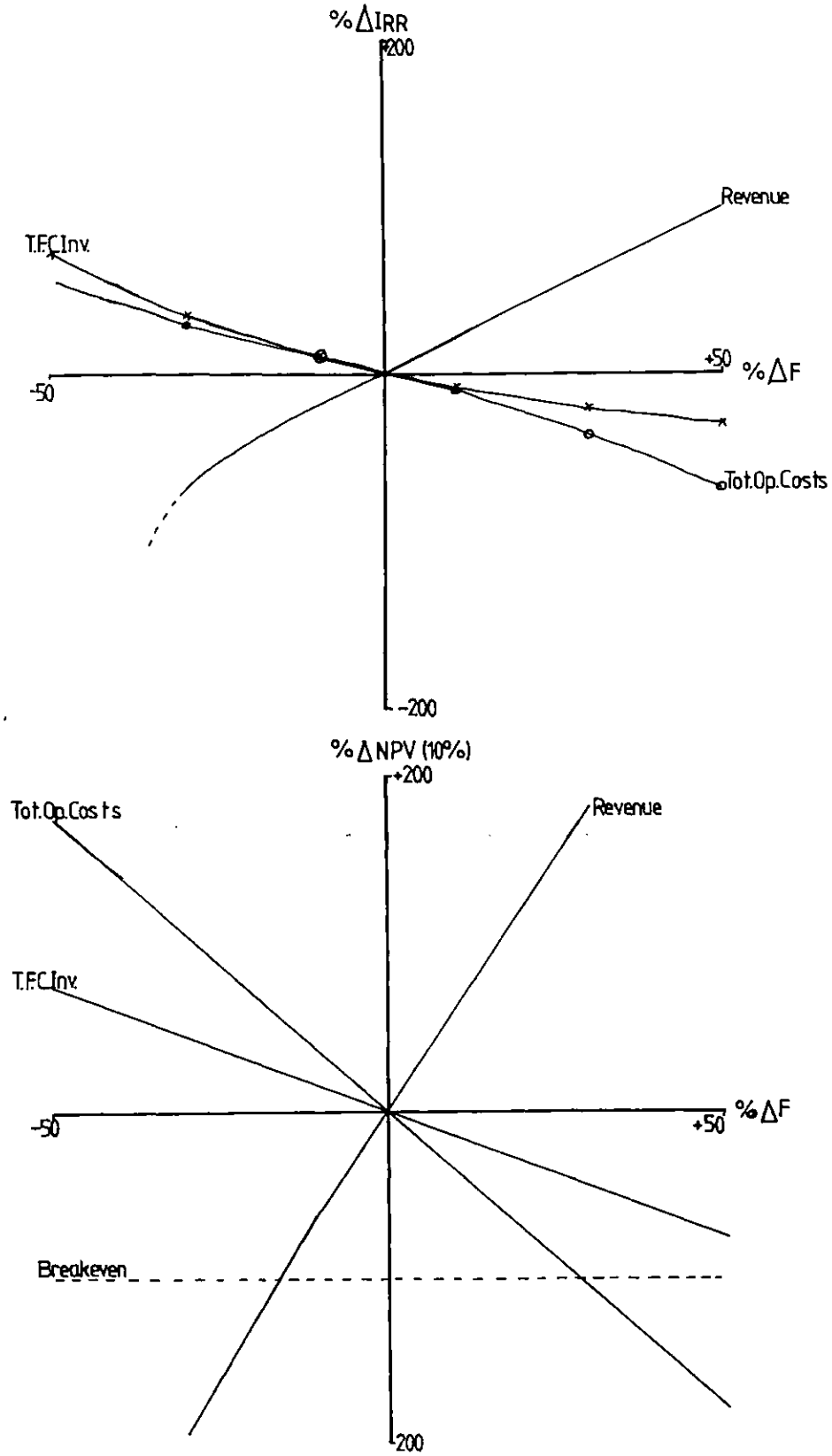


Figure 6.3 Sensitivity of Project Internal Rate of Return and Net Present Value to Project Cash Flows

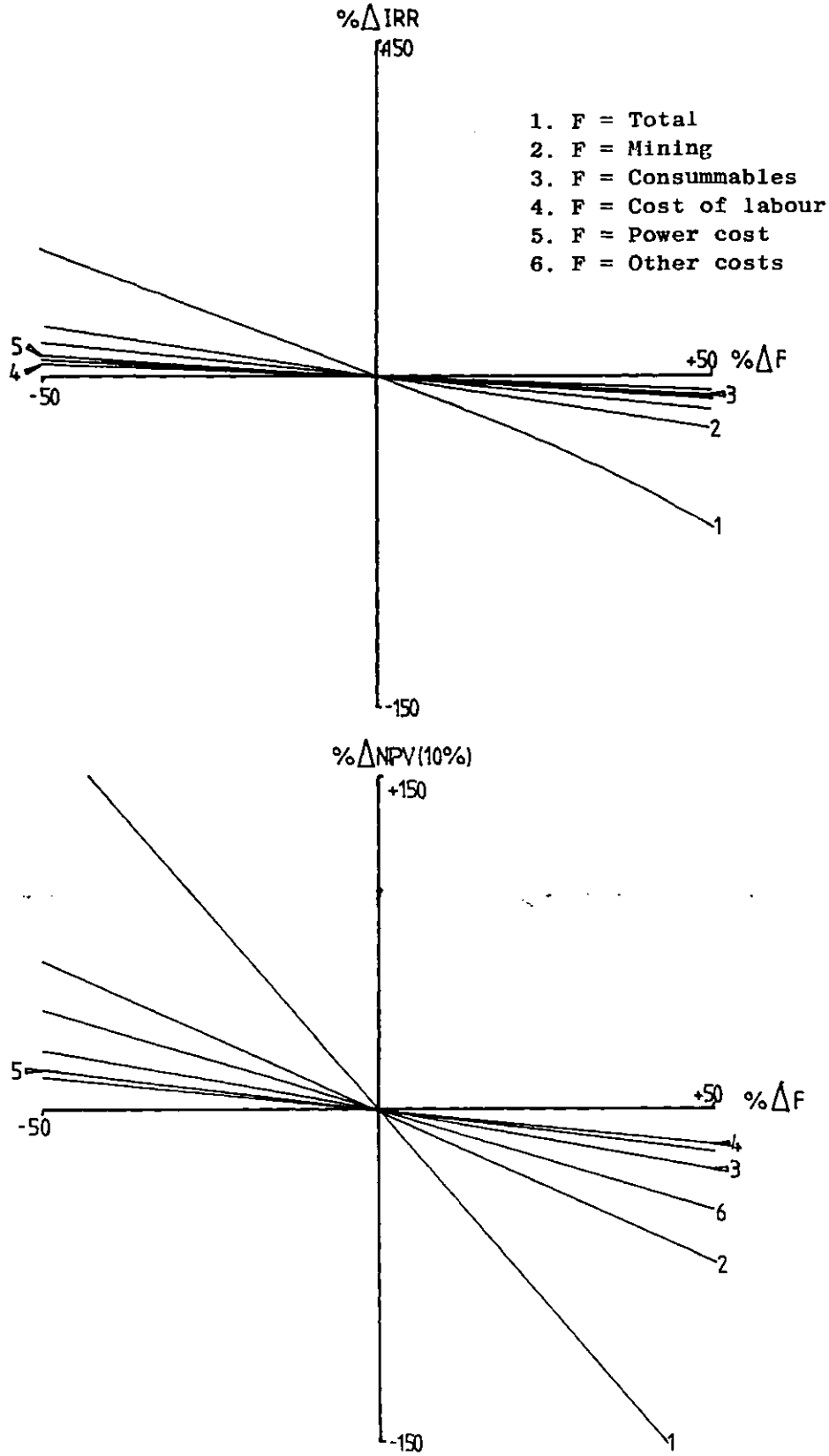


Figure 6.4 Sensitivity of Economic Indicators to Operating Cost Elements

6.2 Technical and Economic Evaluation of Copper Leaching Process

In a similar fashion to the previous case study, a PLANT/MECON data file was developed to describe a plant producing copper using an agitated leach - counter current decantation - solvent extraction - electrowinning process; the base conditions, summarized in table 6.5, were largely derived from descriptions of the Anamax Twin Buttes operation in Arizona (463,464), though the nominal mining costs and media costs were determined from O'Hara (412) and the USBM (462) respectively. Using the PLANT/MECON system, the summary economic data shown in table 6.6 were calculated; sensitivity studies were carried out on the major cash flow elements and are summarized in figure 6.5. A breakdown of the project operating costs is shown in table 6.7.

The project can again be seen to appear highly profitable at the base conditions, though the drawbacks of the economic model and the inflated copper price should be borne in mind before analysing the detailed economics; assuming that all other conditions remain constant, the apparent breakeven price for the copper electrode products is C$2.13/kg (97c/lb), which is a more probable copper price than the one used. A reduction of 41% in the total project operating costs would achieve a similar breakeven condition. In regard to this, it should be noted that copper oxide agitated leaching plants are usually associated with higher tonnage, sulphide flotation plants and are often considered as a by-product recovery operation, processing material normally regarded as waste; it is therefore likely that the proportion of the mining costs attributed to the leach plant would be considerably smaller and may possibly be taken as zero. If no mining costs were to be charged, the economic analysis would be more realistic; this situation would result in an increase of 54% in the net present value. This would be similar to evaluating the incremental benefits to the project of building the leach plant.

A corrollary to the effect of possible metal prices on the project economics is that if the leach plant recovery was to fall from 90% to 69%, then the project would be financially unattractive even at the inflated metal price; obviously if the metal price was lower then the economics would be even more sensitive to overall recovery and ore grade. As the leaching section also accounts for most of the plant acid consumption, which comprises 27% of the plant operating costs,

Location : Montreal

M + S (M.M.) Index at Start-up : 800

Project Lifetime : 15 years Average Annual Inflation : 10%

Taxation : 52% annually 100% capital allowances

Operating Schedule : 49 weeks annually

Mining costs/tonne : $3.85

Feed Tonneage : 10000 tpd (7 days per week) (1.27% recoverable copper)

Plant Product 99.99% Copper electrodes : 800 tonnes/week at $2750 tonne^{-1}

Cost of consummables :		
Crushing steel	$1.55/kg	(Consumption 0.031 kg/ crusher kWh)
Grinding steel	$0.57/kg	(Consumption 0.08 kg/ mill kWh)
Sulphuric acid	$0.03/kg	(Consumption 114.0 kg/ tonne feed)
Flocculant	$4.08/kg	(Consumption 0.2 kg/ tonne feed)
LIX reagent	$11.3/kg	(Consumption 0.066 kg/ tonne feed)
Power	$0.03/kWh	
Operating Labour	$10.0/hr	(basic pay)

Crushing Section

3 operators/shift : 18 shifts/week

1 primary gyratory crusher 1.49 m x 1.02 m : 336 kW motor

5 storage bins : 3,000 tonnes each

1 secondary cone crusher 2.13 m 380 kW motor

2 secondary screens (double-deck) 2.44 m x 6.1 m

2 tertiary cone crushers 2.13 m 2 x 380 kW motors

+ various connecting conveyors

5 fine ore storage bins : 3000 tonnes each

Grinding Section

2 operators/shift : 21 shifts/week

4 ball mills 3.81 m x 9.14 m : 4 x 2235 kW motors

Leaching Section

1 operator/shift : 21 shifts/week

5 agitated leach tanks 620 m^3

4 CCD thickeners 122 m diameter
3 pH adjustment tanks 155 m^3
2 thickeners 122 m diameter
1 tailings dam

Table 6.5 : Copper Leach Plant Base Conditions
(all prices in Canadian dollars)
(Continued)

SX - EW Section

10 operators/shift 21 shifts/week

12 mixers 75.6 cu.m volume

12 settlers 891.5 cu.m. volume

870,000 kg solvent inventory

Electro-winning tankhouse capacity : 630 tonnes Cu electrode/week

operating voltage 2 V

current density 162 A/m^2

Table 6.5 (Continued)

Capital Costs

Battery limit material cost	$50.57 m
Battery limit labour hours	$ 0.779 m
Total fixed capital investment	$143.4 m
Working Capital	$14.6 m

Operating Costs (1st operating year)

Consummables cost	$21.92 m
Cost of labour	$ 7.54 m
Power cost	$ 4.84 m
Other op costs	$18.71 m
	$53.01 m
Mining costs	$16.00 m
Total project operating costs	$69.01 m
Revenue (1st operating year)	$130.5 m

Taxation no taxation payments in first year due to capital allowances

Net cash flow (1st operating year) + $61.5 m

Project internal rate of return	36.88% (24.43% in real terms)
Net present value (21.0%)	£100.0 m
Pay back period (undiscounted)	2.6 years

Table 6.6 : Copper leach project economics

(all costs in Canadian dollars)

	$m	% of total
Crushing : Power	0.46	0.9
Liners	0.53	1.0
Grinding : Power	2.80	5.3
Media	3.50	6.6
Leach : Power	0.03	0.1
Acid + Floc	15.22	28.7
SX - EW : Power	1.55	2.9
Solvents	2.67	5.0
Total Power	4.84	9.1
Total Consummables	21.92	41.3
Operating Labour Pay	1.55	2.9
Maintenance Labour Pay	4.73	8.9
Supervision cost	1.26	2.4
Operating supplies	0.90	1.7
Maintenance supplies	3.15	5.9
Administration	4.52	8.5
Payroll burden	3.02	5.7
Other costs	7.12	13.4
	53.01	100.0

Table 6.7 : Breakdown of operating costs
(all in Canadian dollars)

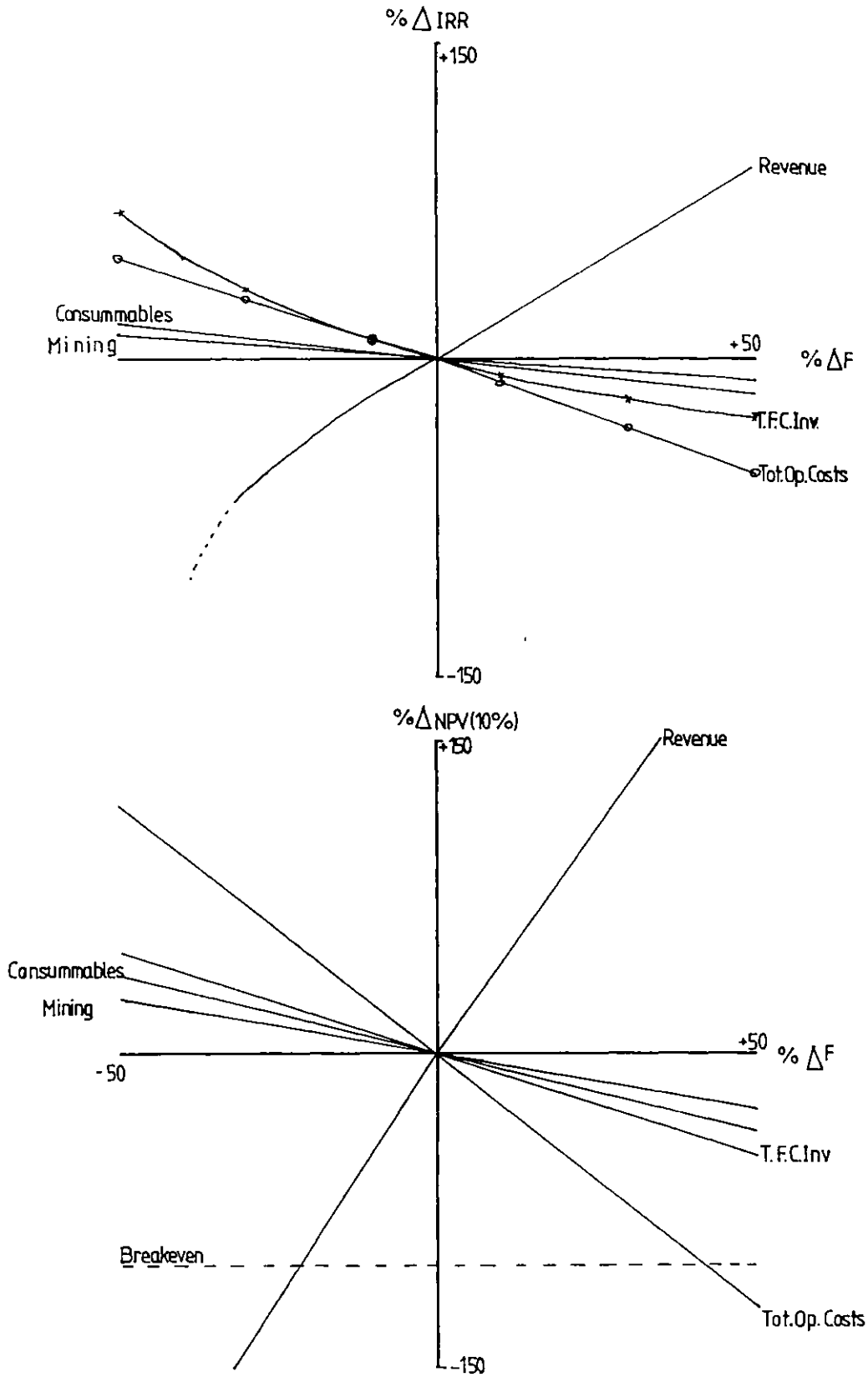


Figure 6.5 Sensitivity of Leach Process Economics

it can be seen that this section is the most crucial to the success of the project and therefore the technical studies should be most concentrated here.

Therefore, to study the technical factors involved with the leach section and to illustrate the use of process models of the leach process in process design, a number of simulation runs were carried out using the LEACH program described in Chapter 5.3. A complete modelling study of the process plant was not carried out, as the use of the comminution and sizing models is illustrated in Chapter 6.3 and the unreliability of the solvent extraction module precluded an evaluation of the downstream processing of the leach liquor and waste solids.

The process model used in the LEACH program has been described in Chapter 5.1 and is based on second-order polynomials where the model variables considered are particle size, solvent concentration, residence time, pulp density and temperature. Although the equation forms have been used extensively in the analysis of uranium leaching (309), there was no data available which could be used to illustrate their application in copper oxide leaching. It was therefore necessary to develop a set of regression constants from the experimental leaching of crysocolla ore (303) using the MINITAB package (88); the equations developed are described in table 6.8(a) which describes the base conditions used in the analysis. In the original testwork temperature and pulp density were not considered as variables and therefore are not included in the equations nor are regarded as parameters to be studied. Ten consecutive leach tanks were modelled so as to illustrate the progress of the extraction, although such a large number would not normally be used. Acid concentrations were set to be the same in each leach tank so that the recovery in each tank may be compared; the LEACH module in ASPEN PLUS, the same as used in this study, allows a different acid concentration in each tank, either controlled by the designer or according to the control of the extraction reactions. The results of the modelling using the base conditions are illustrated in table 6.8 (b); these show that recovery is most readily achieved in the first five tanks and levels off thereafter. A facility not included in the present module but which may prove useful would be to calculate the extra costs incurred by each consecutive tank and to

Number of tanks 10

Tank volume 600 m^3 (100% active)

Fraction of flow bypassing tank 0.05

Solids flowrate 320 kg/sec (1.0% Cu)

Solids density 2700 kg/m^3

Fluid flowrate 480 kg/sec (15 g/l acid)

Average residence time per tank 0.28 hrs

Acid concentration in each tank controlled at 30 g/l

% Cu extracted = -2.82 + CACID + 66.1 TIME -247.0.SIZ
+ 1.66 CACID.TIME -5.43.CACID.SIZ -156.0.TIME.SIZ
-0.0184.CACID.CACID -22,0.TIME.TIME + 807.0.SIZ.SIZ

where CACID = 5.92 + 0.629.CACID -17.1.TIME -0.275.CACIDI.TIME
+ 0.0012.CACID.CACID. + 11.4.TIME.TIME

and CACIDI = acid concentration inlet stream (g/l)
CACID = acid concentration tank (g/l)
TIME = mean residence time (hrs)
SIZ = mean particle size (mm)

Table 6.8(a): LEACH program base conditions

Feed grade	: 1.0% Cu	Tail grade	: 0.0371% Cu
Solution [Cu]in	: 0.0 g/l	Solution [Cu]out	: 7.42 g/l
Solution [acid]in	: 15.0 g/l	Solution [acid]out	: 30.0 g/l

Copper extraction 96.29%

Acid make-up necessary : 7.6855 kg/s (2.49 kg acid/kg metal)
(24.03 kg acid/tonne ore)

Installation cost $727 731.5

Annual reagent cost $11 392 565.0 (at $0.03/kg)

Table 6.8(b): LEACH program base results

compare these with the extra recoveries/revenues achieved.

The first variable in the model equations to be studied was the acid concentration; the sensitivity of the copper recovery and reagent cost to this variable are shown in Figure 6.7. Figure 6.7 shows that the total recovery is more sensitive to decreases in the controlled concentration than to increases; this indicates that at 30 g/l the extraction rate is approaching a maximum and, therefore, also an optimum. Figure 6.7 also illustrates how much more sensitive the reagent cost is to this variable than the total recovery. Using this information it should be possible to optimize the controlled acid concentration in each tank; if however the acid concentration is different in each tank this optimization would be more difficult to achieve.

The second model variable studied was the pulp residence time in each tank. The results shown in Figure 6.8 illustrate the response of the model to changes in the tank volume; the sensitivity of the recovery is the same if the residence time is changed due to a different feed flowrate, though obviously this would have no effect on the capital cost and slightly less effect on the operating costs. In a similar fashion to the changes in acid concentration, total recovery is more sensitive to decreases in tank volume than to increases; that this indicates that a maximum in the copper extraction is being approached is reinforced by the greater sensitivity to changes, of the copper recovery after 5 tanks. Again, the knowledge of how the total recovery changes with the tank residence time can be combined with the results of the changes in capital and operating costs to optimize circuit design.

Although the third model variable, particle size, was not studied using sensitivity analysis, the difference in copper recovery from each size range is illustrated in Figure 6.9. Whilst recovery from -150 micron material is satisfactory, above this size the copper extraction rate drops quickly; if the curve is extrapolated a recovery of only 64% is achieved from 300 micron particles.

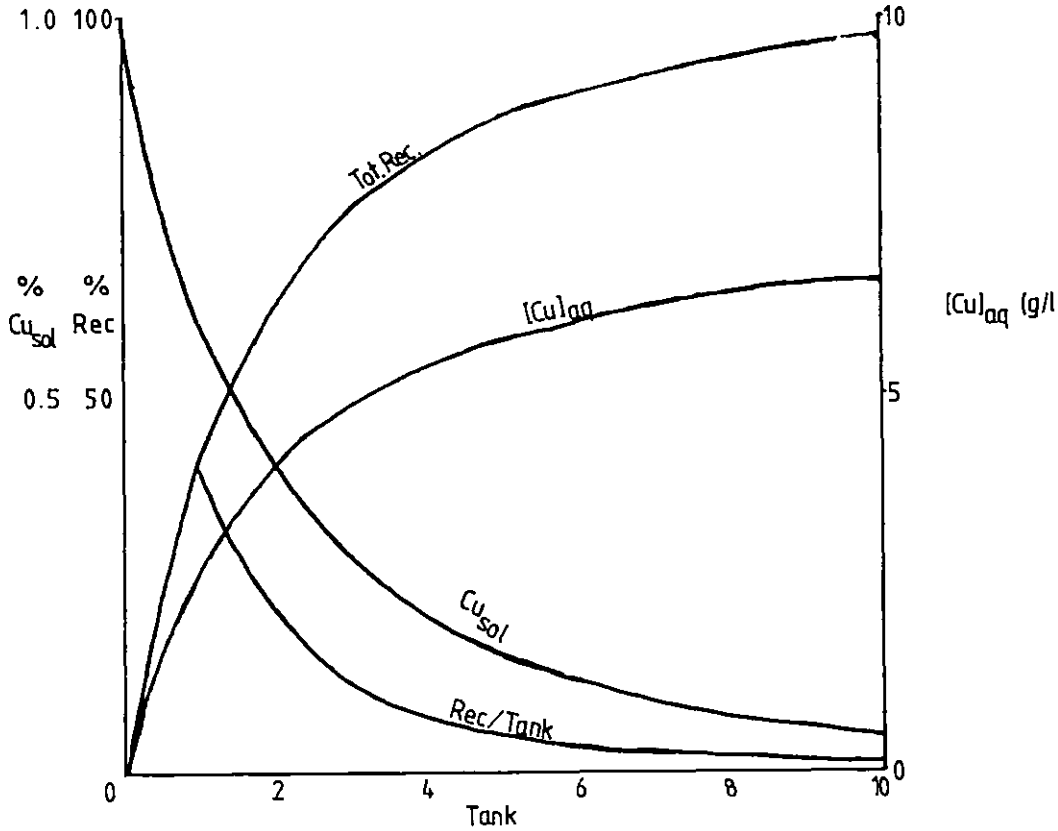


Figure 6.6 LEACH base results

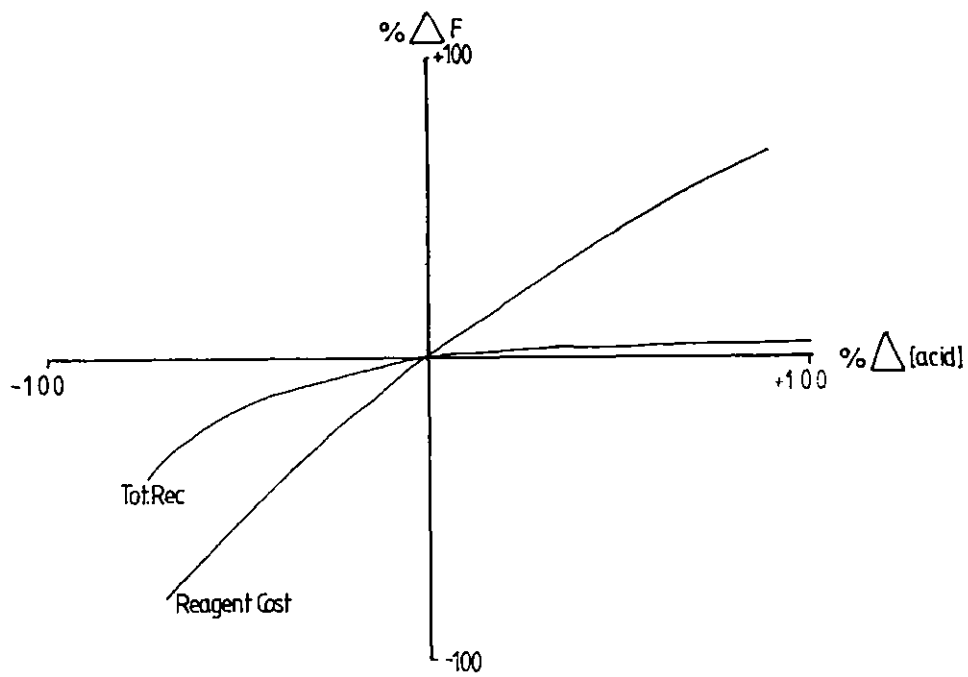


Figure 6.7 Sensitivity of copper extraction and reagent costs to acid concentration

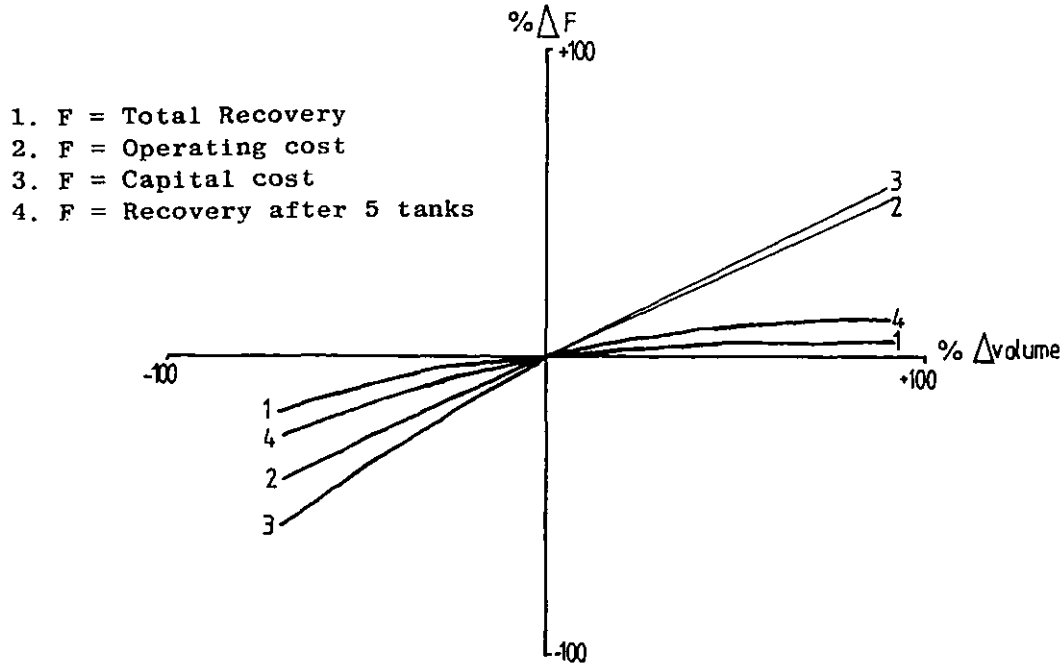


Figure 6.8 Sensitivity of LEACH results to changes in tank volume

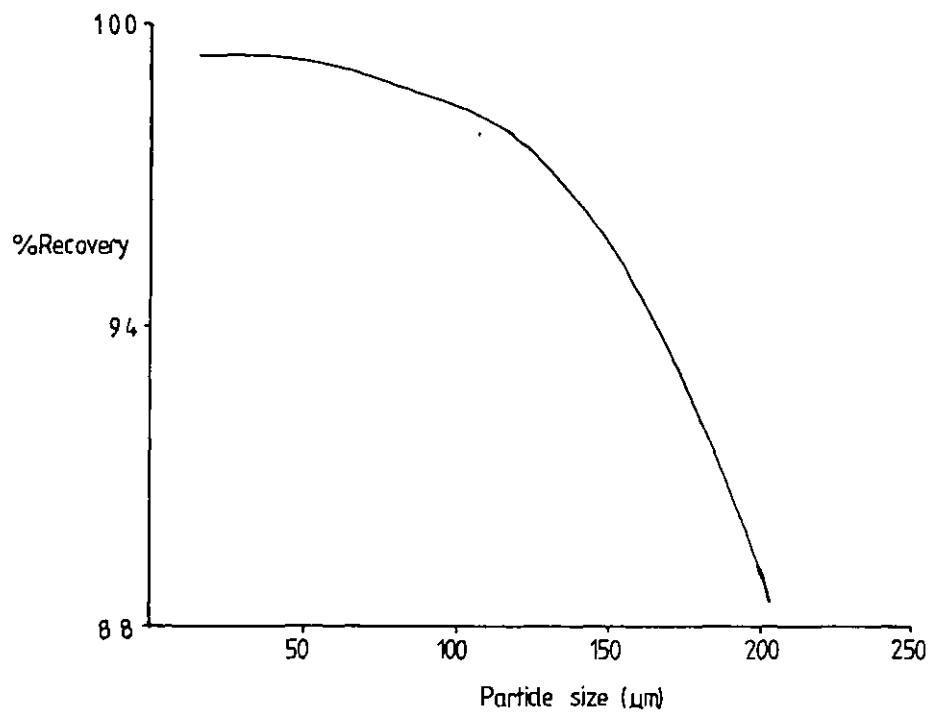


Figure 6.9 Change in copper recovery with particle size

6.3 Technical and Economic Evaluation of a Flotation Plant

In order to demonstrate the use of the mass balance and cost models of comminution, size classification and flotation processes, a case study was carried out to evaluate the effects of varying process variables on the performance of a hypothetical flotation plant using the programs described in Chapter 5.3.

Due to the lack of a module to adjust the size distribution description parameters of a particle population between coarse run-of-mine ore to finely ground flotation products, it is presently more inefficient to simulate a complete process plant than to simulate it in separate sections according to the size of material handled. There is also a need for a module to describe the liberation of minerals according to the size distribution; this deficiency also leads to the handling of sections separately, the distinction in this instance being made between comminution processes, where particles are treated as homogeneous in composition, and separation processes, where particle composition is of primary importance. Due to this latter problem, it is difficult to fully evaluate the implications of changes in the comminution circuit, as their effects on the performance of the flotation operation cannot be measured. Another consequence of this compartmentization of the process plant is that it becomes more difficult to calculate a complete water mass balance and to include models of the solid-liquid separation operations.

Accordingly, the hypothetical flotation plant was split into three sections, crushing, grinding and flotation, and each was studied in isolation from the other, thus reducing the advantages accruing from the use of process simulation in the analysis. The flowsheet of each section is constrained by the structure of the relevant simulation program.

6.3.1 Crushing Circuit

The flowsheet of the crushing circuit studied is shown in Figure 6.10 and the base operating conditions are described in table 6.9; the feed to the circuit is considered to be supplied from a primary gyratory crusher located separately and not included in this simulation. In the costing sections of the analysis, the cost of materials handling equipment, such as conveyor belts, has not been included as suitable sizing algorithms have not yet been

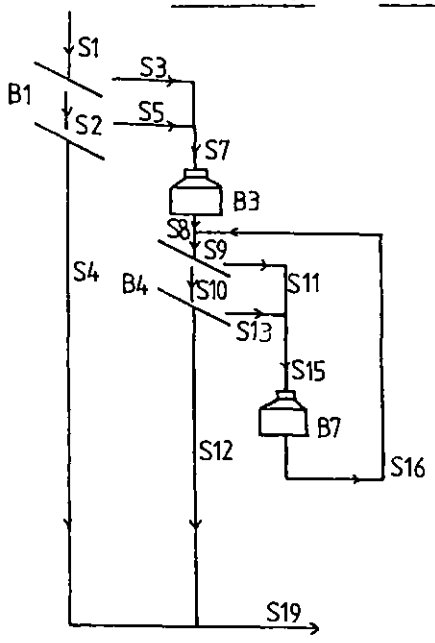
developed; this is a significant drawback as the effects of a process variable in changing stream flowrates are not fully accounted for. From the initial mass balance it would appear that whilst the circuit is producing a satisfactory amount of -9 mm product, the design is inefficient in that only a limited proportion (65%) of the material is circulated through the tertiary crushers, resulting in low feed rates and very high reduction ratios in these machines. If the circuit could be more efficiently utilized, a smaller product size could be readily achieved.

Using the mass balance results calculated at the base conditions, sensitivity studies were carried out on the major variables affecting the performance of the secondary crusher and the tertiary vibrating screens. These studies were initially performed on each block in isolation from the rest of the circuit and then as a part of the circuit evaluation. The initial tests were to study the effects of model variables without the resulting changes in performance being disguised by compensating changes in the other parts of the circuit; their results are illustrated in Figures 6.11 and 6.12.

Figure 6.11 shows that there are only two sensitivity curves, one describing the effect of changes in the Bond Work Index and/or the total solids flowrate, and the second describing the reaction to altering the motor size and/or power efficiency. The only variable that alters costs is the crusher motor size. The sensitivity of the performance to the Bond Work Index illustrates the dependence of the model on the empirical Bond relationship, whilst the model is shown to be also very dependent on the more empirical Flavel efficiency factors (79). The effects of changing five different screen model parameters are shown in Figure 6.12 in relation to the D_{50} of the lower screen deck. The wire thickness is shown to be relatively unimportant whilst the apparent insignificance of the screen area and feed flowrate illustrates the initial over-design of the screens due to the low recirculating load. The shape of the response curve for the top screen mesh aperture is caused by its role in scalping coarser particles and hence increasing the effective screening area; decreasing the aperture increases the amount of separation to be achieved on the top deck and decreases the effective area significantly.

Following this preparatory work, similar sensitivity studies were carried out on the complete crushing circuit to study the effects of

Table 6.9 : Crushing Plant Base Conditions



B1: 1 double screen 2.5 m x 6.0 m
0.16 m x 0.01150 m
0.005 m wire thickness
(D_{50} mm 19.81 X 14.99)

B3: 1 2y o/c cone crusher 2.18 m
380.0 kW motor
(D_{50} 22.45 mm)

B4: 2 double screens 2.5 m x 6.0 m
0.016 + 0.01150
0.005 m wire thickness
(D_{50} mm 19.34 + 14.33)

B7: 2 3y c/c cone crushers 2.18 m
380.0 kW motor
(P_{80} 3.32 mm)

Figure 6.10

Circuit Mass Balance

	kg/s	P_{80}	-9 mm
S1	221.0	96 mm	10.3%
S2	56.67	22.5 mm	40.14%
S3	164.33	105 mm	0.01%
S4	33.08	16.5	68.55%
S5	23.59	25.5	0.32%
S7	187.92	102 mm	0.05%
S8	187.92	22.45	22.97%
S9	310.40	21.00	53.37%
S10	270.97	19.5	61.12%
S11	39.43	27 mm	0.11%
S12	187.92	6.5 mm	88.00%
S13	83.05	26.5 mm	0.29%
S15	122.475	27 mm	0.23%
S16	122.475	3.32mm	100.0%
S19	221.0	7.5 mm	85.08%

Process Economics

Total Cap. Inv. $15.877 m

Power cost $0.152 m.p.a.
Steel cost $0.236 m.p.a.
Other costs $2.983 m.p.a.

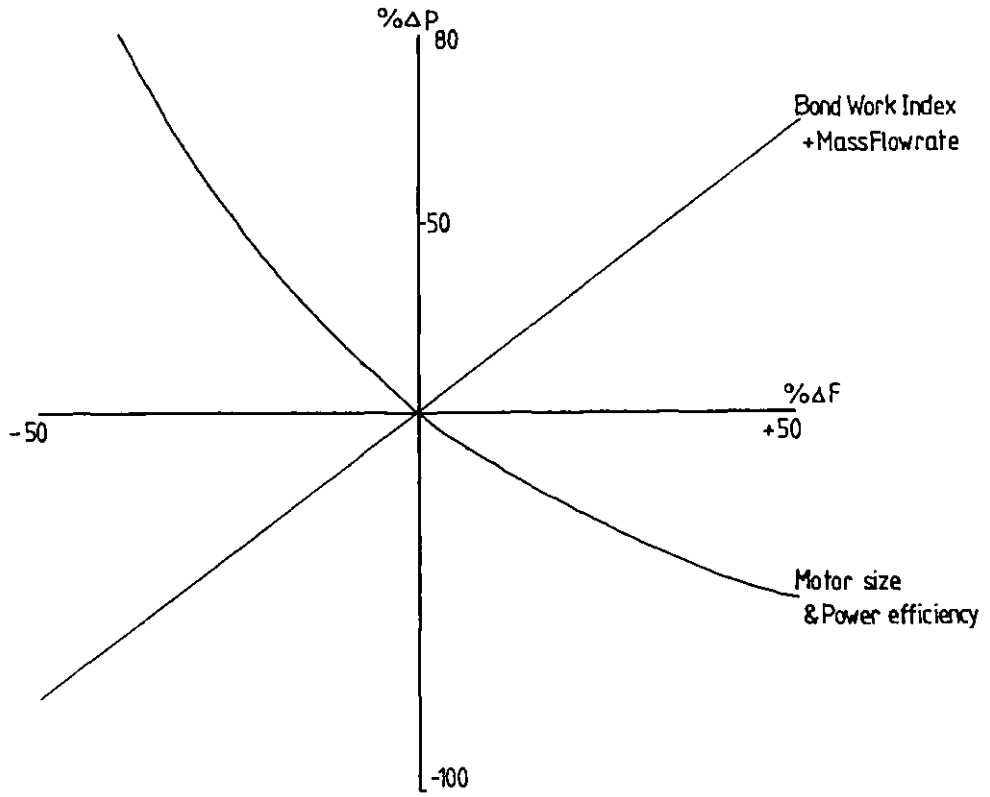


Figure 6.11 Sensitivity of Crusher Performance to model variables

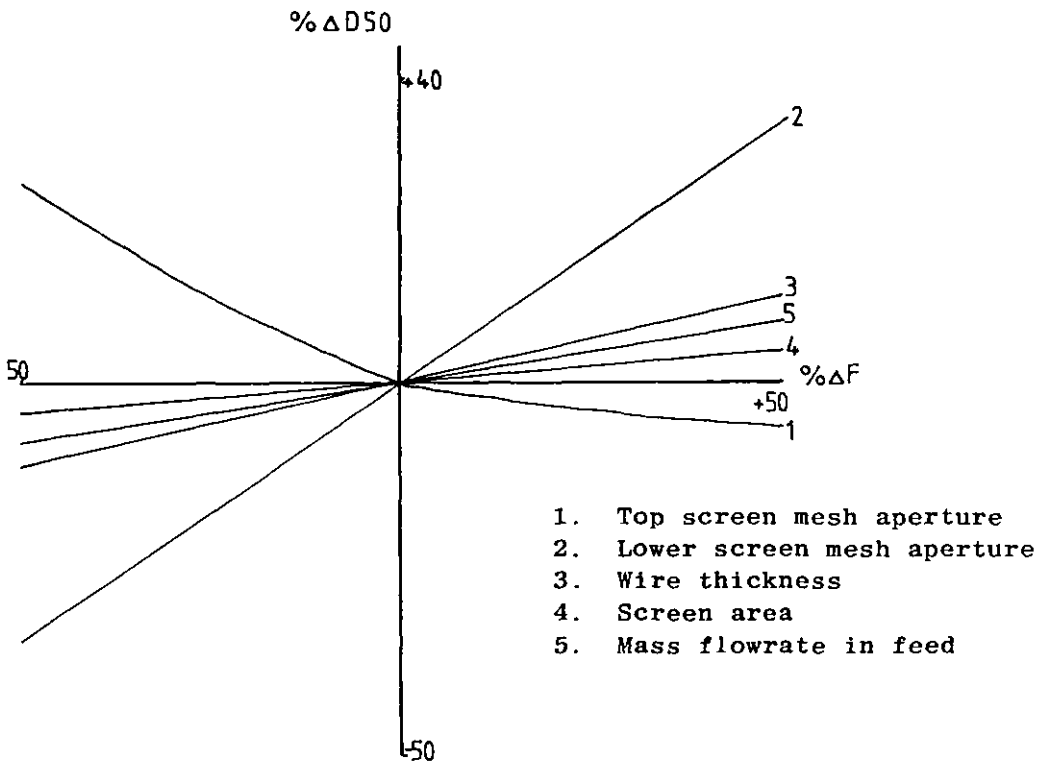


Figure 6.12 Sensitivity of Screen Performance

the more important process variables on the circuit's performance in producing -9 mm material. The results of this work are shown in table 6.10. Of the variables tested the most significant were the screen mesh sizes, though the circuit mass-balance model was not excessively significant even to these, and they had no effect on the cost model of the circuit. The changes in mesh size of the upper screen decks were predicted to have little effect, until the mesh size approached and then became smaller than that of the lower decks; as the mesh size became smaller the upper deck relinquished its scalping role and became the major screening surface, thus significantly reducing the effective screening area. This results in the net cut-size of the screens in closed-circuit with the tertiary crushers being smaller and increasing the re-circulating load; thus these crushers carry out more effective size reductions and the net circuit product is higher in -9 mm material. If it is considered that the lower screen decks are the more effective sizing surfaces and control the product size, then the effects of increasing their mesh size is, predictably, that the overall product size becomes larger; more material is considered of adequate size before crushing begins and less of the secondary crusher product is circulated through the tertiary crushers. When this screen mesh size is decreased by more than 30%, the load on the tertiary crushers increases significantly, with the result of larger crusher product sizes; the screen cut size remains almost constant though, such that there is little change in the overall product size, indicating that the recirculating load is in excess of the screen capacity and that further changes would result in circuit instability.

Whilst changes in the feed flowrate and crusher parameters were sometimes less significant than the changes in the screen variables, they at times produced excessive circulating loads which did not allow the simulation to reach a steady-state mass balance. Even if the convergence acceleration method was replaced by a direct substitution procedure, the most stable type, a satisfactory solution could not be achieved. As the amount of -9 mm material produced by the secondary crusher decreased, the circulating load increased until eventually, above 100% of the secondary crusher product flowrate, the model solution became unstable. Whilst it would appear that the

Variable	% Change in Variable						
	+50%	+30%	+10%	Base	-10%	-30%	-50%
Top Screen Mesh	+1.7	+1.2	+0.5	0	-0.8	+1.1	+14.7
Lower Screen Mesh	-25.8	-16.5	-10.2	0	+9.23	+17.3	+17.4
Bond Work Index	-	-	-0.52	0	+0.7	+0.9	+4.7
Motor	+0.7	+0.6	+0.4	0	-0.4	-	-
Feed Flow	-	+6.6	+1.4	0	-3.0	-6.2	-0.01

Table 6.10 % Change in production of -9 mm material with changes in model parameters

screen area is not a major parameter at or near the base conditions, it became the limiting factor in controlling the amount of material circulated through the tertiary crushers.

From this study of the important process variables it would appear initially that making the lower screen mesh size smaller is the best method of producing a finer product; however more detailed analysis of the results shows that the tertiary crushers are underworked and that more efficient operation would be achieved by increasing the circulating load through them. At the base conditions the secondary crusher is producing enough -9 mm material to make the tertiary crushers largely redundant; therefore efficiency could be improved by making the secondary crusher motor size smaller and/or decreasing the necessary circuit product size. Using the observations made from this initial study, it becomes possible to readily achieve a better circuit design by changing a combination of process variables and/or redefining the overall process design; the route to such an improved design is likely to be a subjective and an iterative procedure, with the design engineer gaining experience of the circuit characteristics through variable adjustment, until an optimal design is produced. The crusher design procedure should be carried out in combination with the design of the grinding circuit as any change in the size of the crusher product will reduce the size reduction to be carried out in the milling section.

With respect to the use of the models in circuit analysis it can be seen that the relative interaction of the unit operation models is more important than the effect of any change in model parameters. The major drawback of the crusher and screen models implemented is that they are only able to simulate a particular situation and cannot be used to automatically design a circuit according to certain control limits. It would be particularly desirable for the user to be able to define the screen cut size and for the model to adjust either the screen area or the mesh size to achieve that cut size; alternatively, the crusher motor rating could be varied in order to control the discharge product size. Such improvements would greatly ease the designer's task and enhance the ability of the unit operation models.

6.3.2 Grinding Circuit

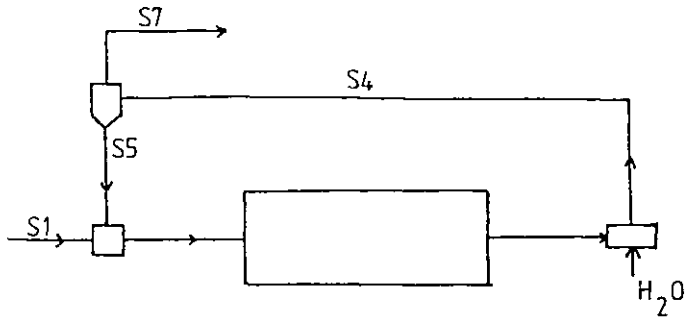
Using the program MILL described in Chapter 5.3 an initial

simulation study of a closed-circuit ball mill-hydrocyclone circuit was carried out; the circuit modelled, the base conditions used in, and the results obtained from this initial study are described in Table 6.11.

As can be seen from the base conditions used, it is necessary to describe the ball mill parameters fully, though the number of hydrocyclones need not be defined; this highlights a potential problem in the use of the hydrocyclone module implemented in ASPEN PLUS, this being that the model does not implicitly include feed flowrate as a variable and that it is assumed that sufficient hydrocyclone capacity is available. Similarly, it is assumed that pump capacity and motor size is adequate to handle the flowrate and maintain the user-defined pressure-drop across the hydrocyclone; a deficiency of the simple equipment sizing and costing routine used is that it can estimate the pump capacity and cost but that it cannot size the motor necessary for a particular duty. These are therefore inherent deficiencies in the modules concerned and should be rectified when ASPEN PLUS sizing and costing modules are implemented; consideration should also be given to means by which a particular hydrocyclone - pump installation may be simulated, including such effects as feed flowrate, rather than using the model in a purely design mode and assuming that any changes in feed flowrate will be handled by the provision of extra hydrocyclone and pump capacity.

Using the simulation base conditions and results, sensitivity studies were carried out to test the effects on the circuit performance of changes made to each of the main variable groups; these studies were made both on the circuit model and on the hydrocyclone and ball mill models.

As can be seen from the description of the hydrocyclone model in Chapter 5, there are only four variables used in the empirical equations describing the solids recovery to the underflow outlet stream; by varying each of these in turn it is possible to test how robust these equations are and where further confirmatory experimental work may be needed. Using the corrected hydrocyclone cut size (D_{50C}) as a measure of the performance, Figure 6.13(a) shows the sensitivity of the hydrocyclone model, whilst Figure 6.13(b) shows the effect of the same parameter changes on the model of the ball mill circuit,



Feed stream (S1)	: 221 kg/sec solids	: 80% - 18.5 mm	: 0.0% - 100 µm
	100 kg/sec water		
Water addition	: 400 kg/sec		
Product stream (S7)	: 221 kg/sec solids	: 80% - 250 µm	: 22.2% - 100 µm
	500 kg/sec water		
Mill product (S4)	: 824 kg/sec solids	: 80% - 620 µm	: 8.5% - 100 µm
	535 kg/sec water		
Hydrocyclone underflow (S5)	: 603 kg/sec solids	: 80% - 770 µm	: 3.4% - 100 µm
	35 kg/sec water		

Ball Mill Model Parameters

Mill Diameter 4 m : Mill length 10 m :

Power constant ($\emptyset_1$) 6.54

Breakage rate function parameters	SF1	0.014
	ALPHA	0.9
	SIZE1	0.0012
Breakage distribution function parameters	BDF1	0.13
	EPSLON	0.3
	GAMMA	0.87
	BE	3.5
Mill residence time definition	RTMEAN	300.0 (sec)
	NMIX	2

Hydrocyclone model parameters

Diameter	0.6 m
Height	1.8 m
Pressure Drop	50 kPa

Table 6.11 Mill circuit model base conditions

using the amount of -100 μm material produced as a comparison. From these it can be seen that the model is most affected by changes in the feed solids fraction and cyclone diameter; the pressure drop across the hydrocyclone and the height have less effect though their relative effects increase considerably as their values are reduced. Whilst it is difficult to judge the model's merits on these results alone it may certainly be concluded that any further study of the model's empirical relationships should especially investigate those concerning the solids content due to their critical effect. Also due to the important effect of hydrocyclone dimensions, it is indicated that care should be taken over making simplifying assumptions about the hydrocyclone geometry such as were made by Ford (16) when developing the modified version of Plitt's model (138) as used in this simulation module. When the limitations of this modified model with respect to feed flowrate are also considered, a return to the original, albeit more complex, version would appear to be worthwhile, as would further testwork to verify the validity of the model equations.

Similar sensitivity studies were carried out on the major ball mill model parameters, the results of which can be seen in Figure 6.14. These show that the total feed flowrate and mean residence time have the most significant effect on milling characteristics, due to their impact on the power input per unit weight of feed; this highlights the need for reliable mill residence time distribution models to be available for common industrial ball mills. One of the major advantages of this type of model formulation over Bond type methods is its inclusion of mass transport effects on milling performance; if however it is impossible to define these properties then the advantages of such models are lost. It can also be seen from these figures that the relationships used to determine mill power and to describe the breakage distribution function also have an important bearing on the model results, and should also be developed further as a matter of some urgency so that the model may be more reliably used. Whilst this more detailed type of model is still believed to possess a number of potential advantages over the use of more traditional design methods, it should be noted that their reliable use in process design is still in the future and a number of important

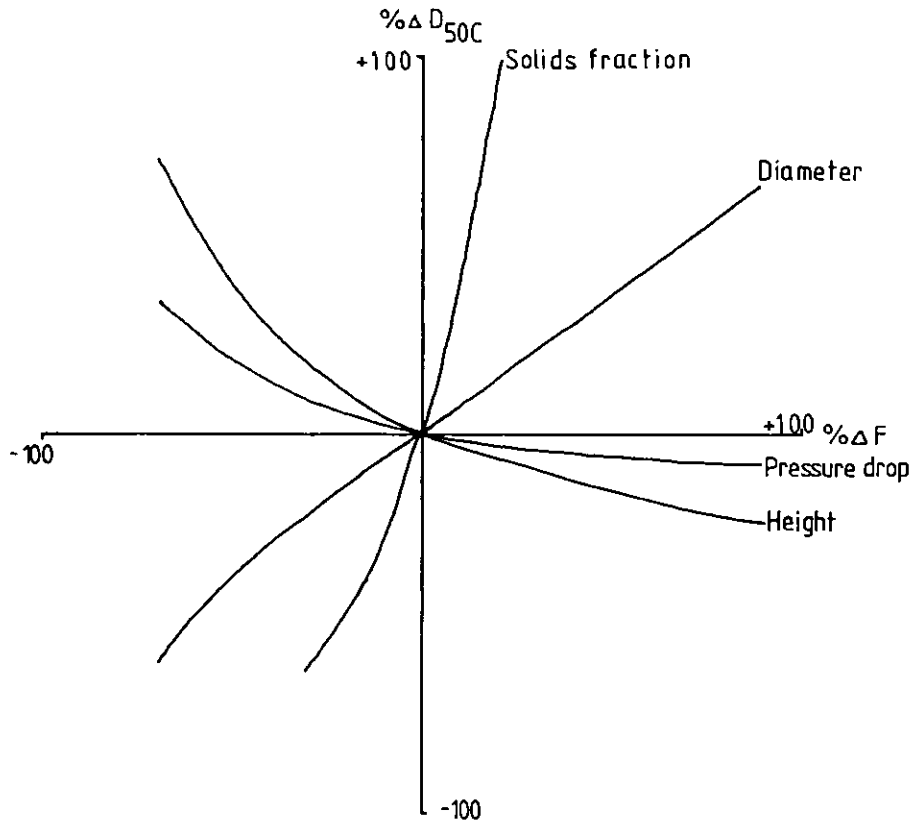


Figure 6.13(a) Effects of changing variables on hydrocyclone performance

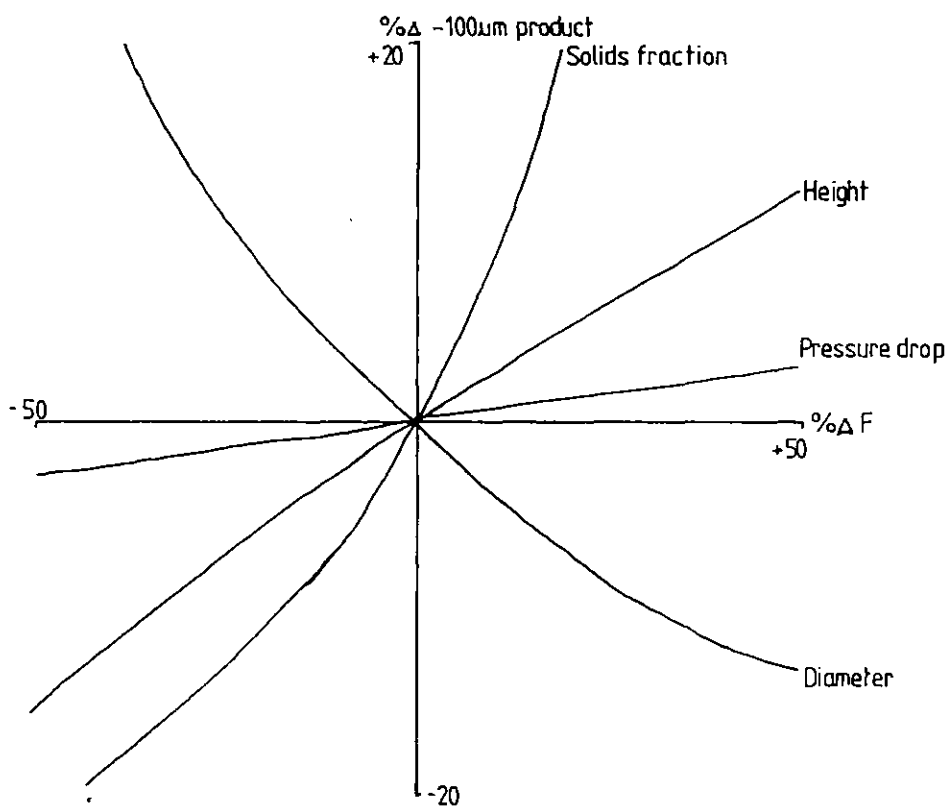


Figure 6.13(b) Effects of hydrocyclone variables on circuit performance

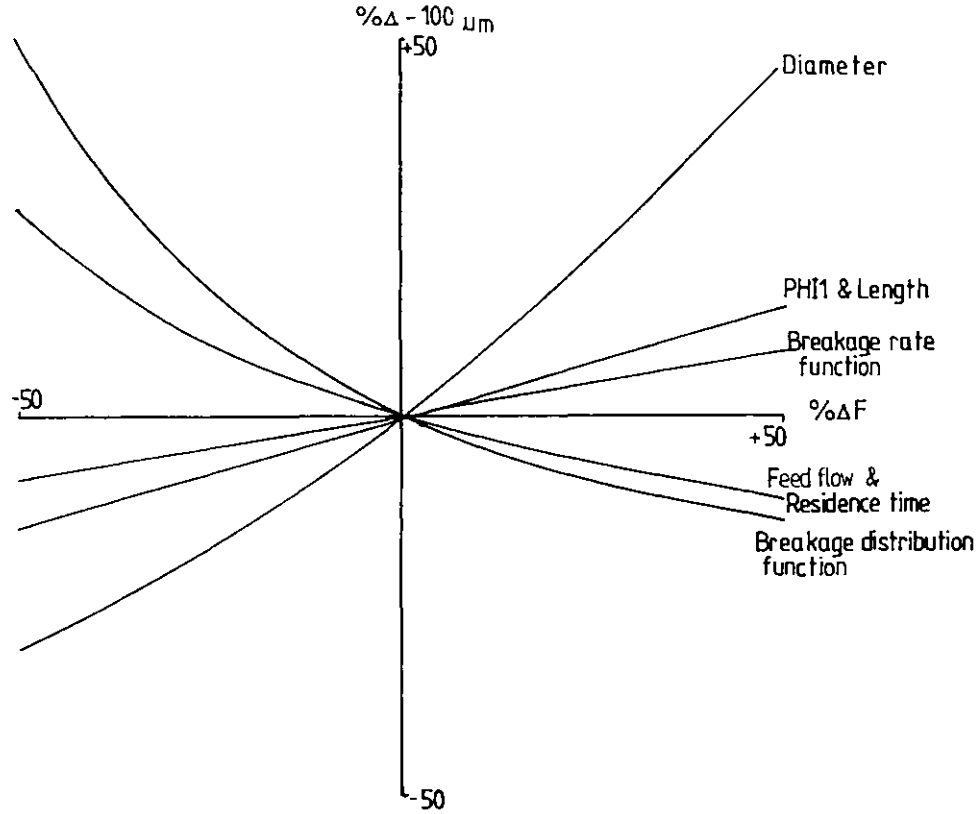


Figure 6.14(a) Effects of changing variables on ball mill model performance

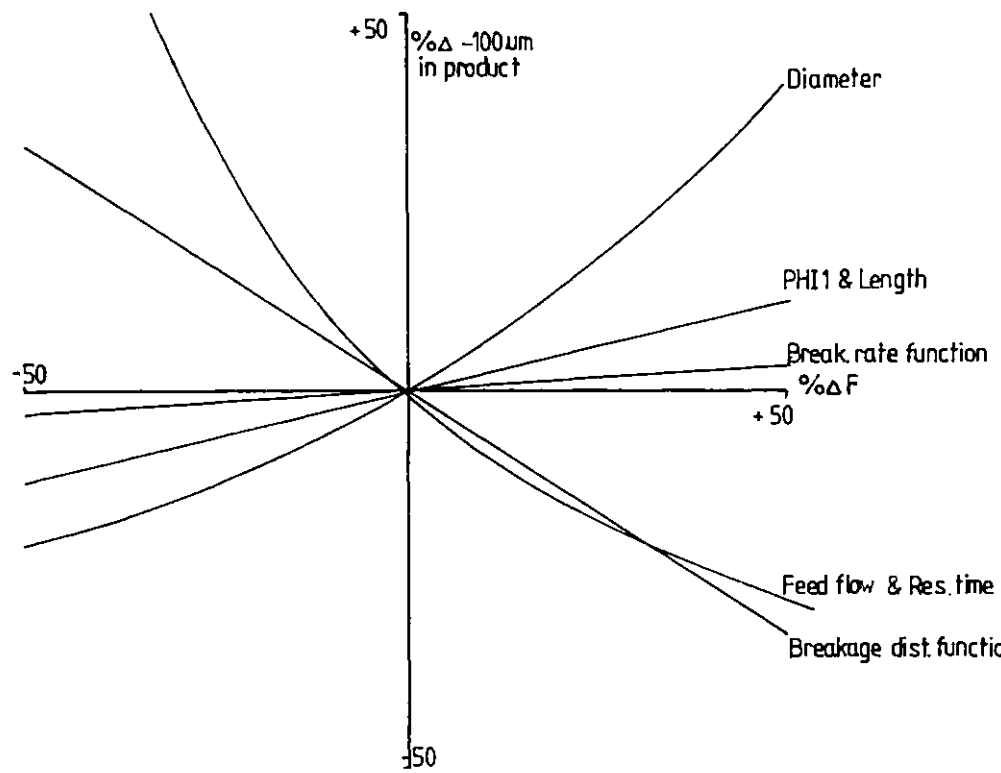


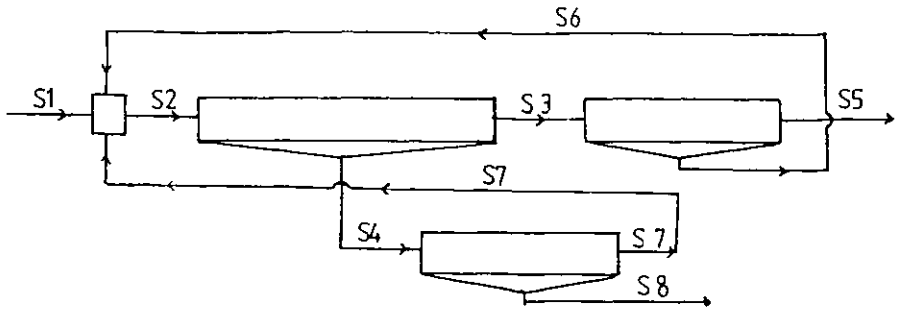
Figure 6.14(b) Effects of ball mill variables on circuit performance

relationships, such as power scale-up, parameter determination and residence time, need to be investigated thoroughly.

6.3.3 Flotation Circuit Analysis

The final case study carried out using the process simulation programs described in Chapter 5.3 was to analyse the effects and relative importance of the major variables in the model of a typical flotation circuit; the circuit and initial starting conditions for the analysis are shown in Table 6.12. The most obvious problem arising from the design and use of the simulation program to be highlighted by this initial analysis, was the definition of flotation reagent consumption. The simulation program used the flotation cost model developed for the PLANT/MECON system, in which reagent useage for a bank of cells is defined by the user in terms of the weight consumed per unit pulp volume of the feed; whilst suitable for a purely cost and economic evaluation algorithm, this calculation method is unsuitable in a simulation environment as the feed flow rate will vary according to the amount of recycled material though it is unlikely that this recycling will affect reagent consumption in an operational flotation plant. Possible alternative methods include using a separate cost block, such as a conditioner, which is positioned before both the head of the flotation bank and the inlet streams mixer, or defining the consumption rate in terms of time rather than flowrate. The problems arising from the sizing and costing material handling equipment such as pumps, occur again in the flotation circuit; these have already been discussed and it appears that a separate sizing and costing block for such equipment is necessary, accessing stream flow information from the simulation though not actually affecting the mass and energy balance calculations.

The variables analysed in the sensitivity studies carried out on this model were the flotation rate constants, pulp flowrate, cell volume, cell tailings pulp density and maximum concentrate flowrate; these variables were chosen to encompass the major model equations and assumptions. The results of these studies are shown figuratively in Figure 6.15, the circuit performance being measured in terms of the overall recovery and concentrate grade. As can be seen from the shapes of the recovery curves it would appear that circuit parameters are close to the optimum for the initial feed conditions,



	S1	S2	S3	S4	S5	S6	S7	S8
Total solids flow (kg/s)	221.0	268.7	211.7	56.9	192.5	19.2	28.6	28.4
Mineral flow (kg/s)	35.35	42.35	13.67	28.6	11.5	2.19	4.8	23.8
Water flow (kg/s)	324.6	394.8	310.9	83.7	282.7	28.2	42.0	41.7

Overall recovery = 67.3%

Rougher Bank
9 x 18 m^3 cells
Tailing pulp density = 1370 kg/m^3
Max. conc. flowrate = 10.0 kg/s
Recovery = 67.6%
Conc. Grade = 50.2%

Scavenger bank
9 x 12 m^3 cells
Tailing pulp density = 1370 kg/m^3
Max. conc. flowrate = 14.0 kg/s
Recovery = 16.0%
Conc. Grade = 11.4%

Cleaner bank
3 x 5 m^3 cells
Tailing pulp density = 1370 kg/m^3
Max. conc. flowrate = 7.0 kg/s
Recovery = 83.2%
Conc. Grade = 83.9%

Total Capital Cost $3.15 m

Ann. Power Cost $0.12 m

Ann. Reagent Cost $1.64 m

Table 6.12 Flotation simulation base conditions

Feed Stream Constituents

Size \ Mineral fraction	1.0	0.8	0.1	0.0
300	4.2 600	4.2 500	73.3 20	38.2 17
250	4.2 800	5.1 700	38.2 50	38.2 30
175	1.9 1333	1.9 1000	1.5 100	1.5 80
80	3.1 1666	1.9 1200	1.5 100	2.0 80

Upper figure = mass flowrate (kg/s)

Lower figure = flotation rate constant ($sec^{-1} \times 10^5$)

Table 6.12 Flotation simulation base conditions.

if it is assumed that maximum mineral recovery is the design objective. However, as can be seen from the lower set of curves, as the circuit recovery increases, the concentrate grade drops; this would normally result in lower revenue per unit weight of concentrate and so circuit design necessitates balancing the value of increased recovery with lower product prices. The value of a simulation system which incorporates costing and cash flow calculations with process modelling, is particularly high in this type of situation.

Whilst the value of the cell tailing pulp density is crucial to obtaining an accurate mass balance for the circuit water flows, it does not appear to have a significant effect, in this situation, on the overall recovery; it would appear, however, that for this particular circuit the defined pulp density of 1350 kg/m^3 is close to a minimum as even a 10% reduction results in negative water flows in the cleaner bank concentrate. The most sensitive model parameter tested, maximum cell concentrate flow, is also the most difficult to measure and is the parameter used to describe the least, or most poorly, understood area in flotation modelling. As was described in Chapter 2.4, the effects of the cell concentrate froth and the passage of material between the pulp and the froth are the most controversial in the topic; several empirical methods for describing the pulp action, including that used in this model, have been proposed, though none have been found satisfactory. This is, without doubt, the area of flotation modelling most in need of further investigation; with respect to the model presently implemented on ASPEN PLUS, work is needed on testing the use of the relationship linking the defined maximum cell concentrate flow and the actual flow, characterizing actual flotation cells according to this parameter and improving laboratory flotation testing procedures so as to more accurately simulate the industrial flotation cell and remove the need for these empirical scale-up relationships. Another model assumption tested using the simulation program was that of perfect mixing in each cell; by increasing the number of cells whilst maintaining the total bank volume, it was possible to measure the effect of increasing the amount of mixing in the cell. However even increasing the number of mixers eight fold, only resulted in a change of 1% in the overall circuit

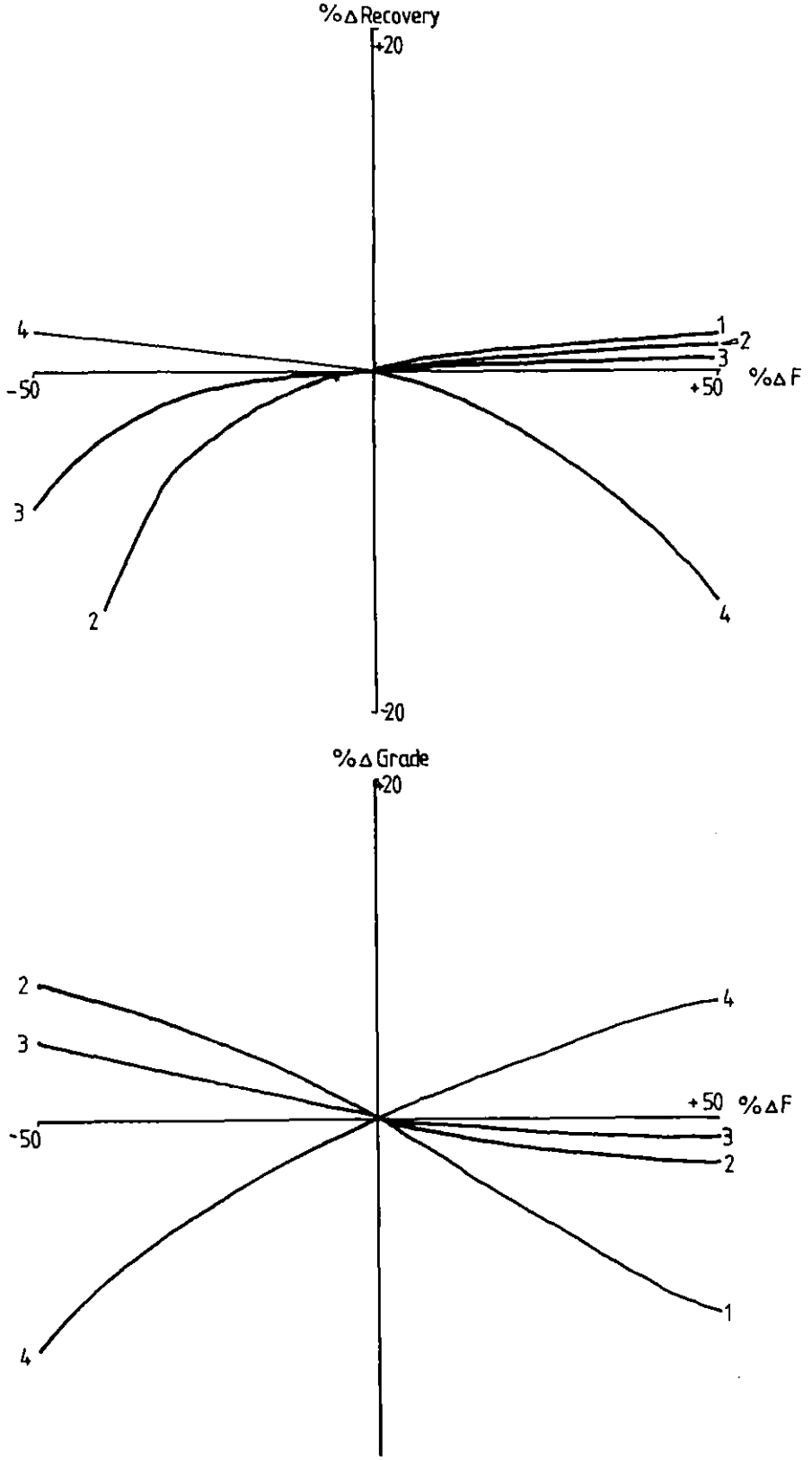


Figure 6.15 Effects of model variable changes on flotation circuit performance

Key F = 1. Tail Pulp density
2. Maximum cell concentrate flow
3. Rate constant/cell volume
4. Feed pulp flow

recovery. It would appear likely therefore that the assumption of perfect mixing in each cell has only a minimal effect on circuit performance; for cases where it is believed that mixing could have a significant effect, it is possible to improve the existing module, with a model based on Cholette and Cloutier's model of an agitated tank as used in the leaching module.

Other than the modelling of the froth effect, it would appear that the only major deficiency in the present model is inability to measure the effect of parameters such as pulp density on the flotation rate constant and therefore to see how the rate constant varies around the circuit, thus requiring the definition of rate constants and the assumption that these are measured under appropriate test conditions for the actual cell environment.

6.4 Simulations using the ASPEN PLUS system

As was described in Chapter 5.1 several additions of stream structures and unit operation modules were made to the ASPEN PLUS system to extend and improve its use in simulating minerals extraction processes. To ensure correct implementation of the eight individual modules, the simulation outputs of each were checked against hand calculated model solutions. The only major drawback of these unit operation modules was the lack of capital and operating cost modules to augment the mass balance calculations and provide an integrated process analysis capability. However development of such modules should not prove too arduous a task in any further work carried out in this area, as the only work these need to do is to obtain the equipment size information, either from user input or mass balance modules, use this data in cost correlations, and provide the calculated results to the overall costing and economic evaluation subsystem.

Whilst there was only restricted opportunity to use the modules in flowsheet simulations, this was sufficient to gain experience in this use of ASPEN PLUS in order to determine possible useful modifications, further applications and restrictions; recommendations for the future development of the system for use in the minerals industry are made in Chapter 7 of this thesis. The system has been widely used in various industries for process analysis and so it may be assumed that the general use of the system has been adequately tested and proven.

Whilst this use of the new modules with ASPEN PLUS was valuable for obtaining general experience of the system, it was also useful for testing the robustness of the modules in process simulation.

As became apparent with the instability of the solvent extraction model, the modules used need to be able to handle input stream data considerably different from what may be considered normal operating conditions. This is due to the iterative nature of the flowsheet solution methods used to obtain an overall mass balance. Initial guesses at stream values are made, these are used in the unit operation modules to calculate a process model solution which is then compared to the estimated solution values; if the two are not approximately equal, the calculated values are used to provide new starting values and the procedure repeated until a stable solution is reached. A

drawback of this solutiom method is that if the estimated input stream values are outside the limitations of a unit operation model's application, incorrect, often absurd, solutions may be obtained resulting in instability in the calculation procedure. Whilst initial guesses at stream values may be provided by the ASPEN PLUS user, it is preferable for the unit operation modules to be more rigorously designed so as to be able to handle such situations; this is especially true for empirically based models whose correct application is only within certain limits. A possible method for achieving a more robust design is to include in the model coding, the limits of the module's application; if inlet stream data is outside these limits then a warning could be written to the simulation history file and/or a model solution determined by assuming the inlet stream data to be equal to the defined limit values. This latter action should result in a more rapidly converging process solution.

Despite the problems occuring with the solvent extraction module the other unit operation modules were all tested in a variety of process situations, though due to the lack of data to be compared against and the lack of time for a systematic testing programme, this was of a subjective nature, resulting only in the proving of the module's implementation and general experience in the use of the ASPEN PLUS system.

6.5 Summary and Discussion

Despite the restricted availability of the ASPEN PLUS system it is felt that significant progress has been made in developing the methodology and demonstrating its potential application and usefulness. A major aim of any future development of the methodology and the ASPEN PLUS system must be to implement suitable equipment sizing and costing routines so that the extensive process simulation capabilities may be integrated with the economic evaluation subsystem to provide a powerful tool for process analysis. A possible problem in this area is how the process economic studies may be included as a part of the overall project economics so that more reliable estimates of the mining and other project costs may be provided; as was seen in the two economic case studies the mining costs alone can be highly significant in relation to the total operating costs. As this initial system development work is completed it is essential that more reliable and exhaustive testwork is carried out on the use of the models and methodology techniques; this should encompass the absolute and relative accuracies of individual unit operation models, and the accuracy and robustness of the overall process models based on these.

One important aspect of process economic evaluations to be highlighted in the first two case studies is the proportion of both capital and operating costs to be estimated using cost factors. Whilst stressing the need for reliable cost factors to be available, perhaps the most important conclusion to be drawn is that these only define the relative importance of cost areas and that it is essential for the basic costs to which these are applied to be accurately estimated; no matter how detailed or accurate a cost factor is, if the costs to which it is applied are inaccurate, then it is impossible to obtain a dependable cost estimate.

Each of the overall process economic evaluations were highly sensitive to the revenue accruing from the sale of products and it may be assumed that similar evaluations will also generally be most sensitive to the size of incoming cash flows. In a process simulation context therefore the most important unit operations are those in which the product is initially extracted from the ore and, to a lesser extent, those which have a large effect on product quality. The former category includes leaching and flotation, in which recovery and grade are largely a function of model variables; the second category includes

concentrate dryers and solvent extraction-electrowinning circuits, in the models of which either controls are applied to, or assumptions are made about, the actual product composition and in which overall recovery is generally not affected. Whilst it is therefore advisable to give detailed attention to the recovery unit operations, it is also very important to reliably model upstream unit operations, such as comminution, which directly affect their performance characteristics. However, not only is it important to model the unit operations, but it it is also necessary for the stream data linking the models to properly represent the real-life material flows, especially in terms of particle size distribution and mineral distribution or liberation.

It is still felt that the method used to describe particle classes in discrete distributions is the most accurate, but a major problem lies in the change in the order of magnitude of particle sizes between run-of-mine ore and tailings. It is possible to describe the feed material in terms of a detailed size distribution including all possible size classes of interest; however this would be inefficient computationally, especially at the finishing process stages where a large majority of the defined size classes would have zero flows. A more efficient but potentially less accurate method would be to model each process section using only the most relevant size classes and to derive subsequent size distributions from the product stream data; this is likely to involve fitting the discrete data to a continuous function and then deriving a further discrete distribution from this function, a process which is not only likely to lead to inaccuracies but which also negates many of the advantages of the discrete distribution form.

As was discussed in Chapter 2, comminution processes have not yet been successfully modelled in relation to their effect on mineral distribution or liberation, a major drawback considering the primary objective of such processes. Therefore, before minerals extraction processes can be modelled as a whole rather than in parts, it is essential to develop a technique whereby the effects of liberation may be included in a process model. It is likely that the most suitable method will be to include a module whereby the design engineer may define the mineral distribution within a particle population, on the assumption that this is measured at a size similar to that of the particles.

Whilst further development of the mass balance models will always be of value, it would appear that the existing routines allow a

reasonable analysis to be carried out of circuit performance and especially the relative importance of process variables. From the studies carried out it can be seen that although a parameter may have an important effect on the performance of a unit operation, when a complete circuit is analysed this variable becomes less significant in comparison with the unit operation interactions and overall process performance.

From an overall point of view it would appear that whilst comminution is important in both performance and cost terms, the performance characteristics of crushing circuits would be largely sheltered from the recovery operations; therefore robust, but less detailed models of these operations will generally be adequate from a mass balance aspect but that reliable cost models are essential. If pre-concentration processing, such as heavy media separation, is proposed to be used before the grinding circuits, then it is likely that more detailed modelling may be required.

However grinding circuits are of vital importance in both process simulation and process costing and so there is a need for more detailed, reliable models of the unit operations. This highlights the need for better design procedures to supplement traditional design techniques such as Bond's law and hydrocyclone performance charts. Whilst more detailed models of the grinding process have been proposed these are seen to be dependent on further development and testing in the areas of scale-up relationships and the measurement and definition of breakage functions. From the brief simulation study carried out it would appear that the hydrocyclone model used is fairly robust and so should allow reasonable predictions of circuit performance to be made; however it was felt that the use of the original, more detailed model formulation would be advantageous for studying the effects of cyclone geometry and feed flowrate. It should be noted that this is a highly empirical model and so further experience and calibration is desirable, especially of its modelling of the effect of pulp density.

The flotation model used also proved to be robust, though its accuracy and use in process design should be treated with some caution due to a lack of validating data and experience; this is especially true in its modelling of water flows and froth effects. It is however derived from a family of models including similar detail and of similar structures and so its general form may be assumed to be

reliable. The major factor influencing this model is the measurement of suitable flotation rate constants to model a particular flotation environment, especially in terms of froth effects and pulp conditions; the model used places more stress on the accurate laboratory measurements rather than using further parameters to allow for secondary effects. In terms of capital cost estimation, flotation is perhaps not significant, but it should be remembered to include all flotation equipment in a cost model, such as blowers, paddles, motors and pumps. The cost area where flotation is most significant is reagent consumption and an accurate method for this cost to be included should be developed when the ASPEN PLUS flotation cost block is implemented.

The form of the leach module used allows for a considerable amount of detail to be modelled, though further parameters such as pressure and agitation could be included, and the only possible problem is with the equation forms used, in the case that a particular leach reaction requires a different form. However a generally applicable type was used and should provide considerably more flexibility and potential accuracy than any of the proposed phenomenological models.

Solid-liquid separation operations may be modelled using the new ASPEN PLUS "solidswash" module which has a very simple and robust structure and can be used to simulate any such operation. Whilst some experience is needed in its use in simulating counter-current decantation or similar processes, it is felt that this level of detail is most suitable for modelling solid-liquid separation operations due to their relatively insignificant effect on process mass balances and the available performance data at an initial design stage. Similarly, whilst not small cost items, the relative cost importance of these unit operations in the preceding case studies would indicate that fairly simple costing routines would be appropriate, the major items of importance being capital costs, pumping power and reagent consumption.

Solvent extraction and electrowinning unit operations are also more important in cost terms and robustness of solution, rather than their effect on the overall process mass balance. As was discussed in Chapter 2, very detailed electrowinning models would be required for predictions to be made about electrode quality and so it would appear that the only necessary improvements are in the solvent extraction model solution procedure and adequate cost correlations to be developed, as these are unavailable in the CIMM publication.

As has been stated earlier, whilst the proportion of capital costs estimated using factors is high, it is particularly important for the basic costs to which they are applied to be accurately estimated. As the majority of models implemented include the definition of equipment sizes as model parameters, the most critical capital costing area is the acquisition of suitable equipment cost-size relationships. Another important area is to ensure that all major equipment items are accounted for; a crucial area is that of material handling which encompasses items such as pumps, conveyors, storage and tailings disposal and which can include significant cost items without directly affecting process performance. Major capital cost areas are typically the crushing, grinding and solvent extraction - electrowinning sections and so most importance should be attached to these in the development of suitable cost blocks and installation factors.

In the two economic case studies carried out, the overall profitability was more sensitive to operating costs than capital investment, though it should be noted that mining costs comprised a major part of these. In a similar fashion to capital cost estimation a significant proportion of the total costs are estimated using factors and so again care is needed in their application and the estimation of the basic costs to which they are applied. These are the total fixed capital investment and the operating labour costs. In the estimation of the latter, no unit operation models are used but it is important to include the cost of shift workers off-duty on a working day; for continuous operations this amounts to an extra 33% above the nominal requirements. The operating costs calculated directly from unit operation models are the power consumption and cost of consummables.

In both economic studies the power costs were, surprisingly, relatively low, especially in the comminution circuits where liner and media costs were higher. It is therefore advisable to check the validity of the techniques used, especially with regard to motor ratings, ancillary consumption, supply costs and voltage types. Whilst comminution is predictably the largest power consumer, it is also where costs may be most accurately estimated; therefore sections such as flotation, solid-liquid separation and solvent extraction - electro-winning, where power costs are also high should perhaps be studied more closely. Important aspects include the provision of suitable power-volume relationships for agitated vessels and the estimation of pumping

power requirements.

Consummable costs in both studies were relatively high and so it is important that provision be made for their accurate calculation. All process sections were important and it is important that over simplifications are not made by lumping together different consummable types such as milling media and liners and all flotation reagents. If these costs are defined in unit operation blocks there are possible problems which may arise from how these are measured and defined, such as occured in the flotation simulation study.

Probably the most useful improvements which could be made to the modules implemented on ASPEN PLUS would comprise the inclusion of a design capability in addition to the existing simulation analyses; this could be provided within the framework of the simulation modules and should only require a rearrangement of the present calculation procedures, though more stress would then be placed on the provision of equipment sizing routines and relationships. Possible examples include the design of a crusher to produce a particular product size and the design of hydrocyclones and screens to achieve defined cut sizes.

The major importance of the preceding illustrations of the methodology and its applications is that it has been shown how an integrated mass balance and economic evaluation system may highlight the relative importance of process variables, indicate those areas most sensitive to changes and therefore may be used to determine those areas most in need of further investigation and/or development.

CHAPTER 7 Discussion, Conclusions and Further Work

7.1 Introduction

The complete analysis and evaluation of a minerals extraction plant requires the calculation of a process mass and energy balance, the sizing of process equipment, capital and operating cost estimation and evaluation of project cash flows; this type of analysis, or at least major parts of it, is carried out at all stages of the development of a process, from the research and development stages of a proposed new process to the evaluation of the efficiency of an established, full-scale and operating plant. In the pre-construction phases of a project it has become more difficult to design or choose an optimal process configuration due to the effects of the increasing costs of labour, power and capital,and the multiplicity of process factors to be considered. To date, the ability to engineer efficient processes has been limited by the large effort required for process engineers to evaluate alternative flowsheets; traditionally, manual calculation techniques have been used, allowing only a limited range of process flowsheets and parameter variations to be studied. Many of these traditional methods of plant design and analysis have been found inadequate and the need for increased sophistication has become pronounced.

With the increased availability of computer hardware and the corresponding decrease in the cost of computing, the use of computationally based evaluation procedures is desirable. Due to the highly complex and incompletely understood nature of minerals extraction processes, it is anticipated that it will be a long time, if ever, before a large empirical input to the evaluation procedure becomes unnecessary; computational procedures are only likely to assist the understanding and development of a process through their ability to allow the engineer to study the process in more detail. However the development of such procedures is felt to offer many advantages, especially by increasing the number of process alternatives and variables that can be analysed; by allowing consideration of a complete plant flowsheet rather than isolated sections of it; by giving better emphasis and consideration of cost and economic factors; by improving analysis reliability by reducing the calculation errors and removing the need for restrictive short-cut methods based on unsound generalisations; by

promoting consistency in evaluations and acting as a communications tool between the various groups involved; and by showing explicitly the implications of assumptions made about process behaviour.

Therefore a methodology has been proposed for the evaluation of minerals extraction processes, computer-based and using mathematical models of the constituent unit operations to develop material and energy balances of the complete process, to determine the effects of equipment sizes and to predict capital and operating costs which are then used to estimate project cash flows and profitability. As was suggested earlier, the components of a process evaluation are similar throughout the development and operating stages of a project lifetime, and therefore the methodology is applicable throughout this lifetime. However the amount of information available to develop the mathematical models will increase during this time, improving their reliability and increasing the included detail. Therefore, whilst the computer systems, software and hardware, might be the same, the models used at each stage would almost certainly need to differ.

As the effort required to develop models for each stage, as well as the flowsheeting package or packages to combine them, would be excessive it was necessary to concentrate on one stage in the project development procedure. It was decided that the most advantageous aspect to concentrate on, would be the Type C (33) study level, otherwise known as the preliminary, intermediate or feasibility study type; in this type of study, each potential process has been investigated sufficiently to allow reasonable predictions of its behaviour and a number of alternative configurations will need to be analysed in sufficient detail for a choice to be made on the likely optimal process route; therefore the advantages of a computer-based analysis procedure, supplying both technological and economic evaluations, are likely to be fully utilized in this type of study. Also by selecting a study level between the less and the more detailed study types, there would be a better chance of extrapolating the techniques and systems developed to other study levels.

In Section 7.2 the development of a computer system upon which the feasibility study application of the methodology may be based is discussed; this discussion includes the selection of suitable executive software and suggested necessary further improvements to the chosen

software: methods by which the costing and economic aspects of the methodology may be incorporated into the system and the need for the compilation of process cost data: the implementation and deficiencies of the unit operation mass balance models used to illustrate the application and potential of the methodology: and the need for further models to extend the use of the methodology to more minerals extraction systems.

Although it has now been shown that a computer based method for analysing and evaluating the feasibility of minerals extraction processes can be developed, largely using existing technology, the proposed methodology should be applicable to other levels of study and therefore in section 7.3 other applications of the methodology and its supporting computer software are suggested. Having discussed the methodology techniques and applications, it is necessary to consider, in section 7.4, the places and methods, in and by which these may begin to be introduced, accepted and integrated into industrial minerals extraction process design and analysis.

7.2 Feasibility Study Applications of the Methodology and Computer System

The development of a computer system on which to base the methodology encompassed three distinct areas of study: flowsheeting software, unit operation modelling and cost estimation and economic evaluation. From these studies it was clear that, whilst there was a great deal of work being carried out in each of the respective areas, much of which could be utilized, there was no existing system which could be used to implement the methodology and study its application and implications; it was therefore necessary to start the development of a suitable system. Due to the enormity of carrying out such a project from scratch and the desirability of using proven technology, a great deal of the work entailed adapting existing software and analysis methods. It was found that the areas which were to cause most problems and would require most development effort, were those involved with the actual process modelling rather than the costing and economic evaluation areas, although the output from the latter is perhaps the most significant information and is the largest benefit derived from using a computer-based methodology.

The models and techniques used were chosen as being those most similar to existing design and analysis methods; there was little point in using a mass balance model of a unit operation which required its parameters to be estimated from testwork on an operating plant. Whilst those chosen were likely to be able to maximise the use of the data which could be available during a Type C study, it would appear that there has been little quantitative evaluation of the reliability of the use of any unit operation model for design and the published estimates of the accuracy of factorial capital and operating cost estimation techniques vary considerably; therefore, it is not possible to judge how reliable their combined use is in the evaluation of a complete process flowsheet. Hence, the prime objective of any future development of the methodology and of the flowsheeting system should be to test their use by simulating existing plants and comparing the predicted results with the actual process performances; as there is an almost complete lack of appropriate, published data, it will be necessary to study a plant for which there are existing records of plant performance and laboratory testwork, or to actually collect the data, carrying out testwork on the plant and on the feed raw material. Whilst each plant analysis would involve a considerable effort, validation of the methodology for industrial use is

essential and almost impossible without such work.

7.2.1 Flowsheeting Software

From a survey of those computer simulation systems that have been described in the literature it would appear that, although there were no integrated flowsheeting systems suitable for use in the mineral industry, there is a greater interest in the possible value of process simulation; this is borne out by the number of simulation systems being developed (19-21), both in industrial and academic institutions. However there are many limitations, especially their availability and reliability, to the use of these, and none could claim to approach the sophistication of chemical process engineering simulation packages; it was therefore decided to extend the capabilities of a system from the chemical engineering industry, such that minerals extraction processes could readily be evaluated. As a project developing a system, ASPEN, which used a flexible data structure for describing processes and, more importantly, streams, was approaching completion it was decided to concentrate on the results of that project and, if possible, to use the ASPEN system; eventually the use of the project's commercial successor, ASPEN PLUS, was offered and so development continued using this system. If this had proved impracticable the work would either have used another chemical process system of a less suitable nature or would have concentrated on illustrating the methodology using a more limited set of unit operation models and/or a less flexible executive system.

However through the availability of the ASPEN PLUS system, it was possible to base the methodology on state-of-the-art technology and continue with more far-reaching objectives as the ultimate aim, even if it was impracticable to achieve all of these within the timespan of one research project. It is expected that no comparable system will become available in the foreseeable future and that therefore ASPEN PLUS should be used as the computer system upon which is based the design application of the proposed methodology. However, if it is assumed that the ultimate object of a methodology development programme is a computer system for the evaluation of mineral extraction processes, even excluding the unit operation modelling and economic analysis aspects, there are a number of system changes which could be made to modify ASPEN PLUS for this application; these should, however, be able to take

advantage of the system's modularity and should largely consist of adding and removing modules.

With the rapid increase in the calculation powers of small computers it is likely that a mineral extraction evaluation system will be hosted on a fairly small, by today's standards, but powerful, and centrally located data processor, which can be connected to other processors and accessed through terminals with the computing power of today's personal computers; even though the price of computer hardware is considerably less than it used to be, it is expected to drop further, so that such networks of processor units and terminals are likely to be fairly common in engineering offices in a few years time. Despite this abundance of computing power and because the present ASPEN PLUS system requires considerable computer storage facilities, it would be desirable to reduce the size of the system as much as is possible. The two main areas in which size reductions could be made are the unit operation libraries and the physical and thermodynamic property estimation routines and data banks. The first of these is a simple matter, it only being necessary to remove the unwanted modules and edit the relevant model tables in the System Definition File (see Appendix 1).

The unit operation models which were implemented to illustrate the use of ASPEN PLUS are only used to calculate the mass balance around a process flowsheet and do not call on any physical property estimation routines; all of the necessary properties for solids are carried in the component attribute locations of the stream structure. The property estimation sections comprise about a half of the system and so a reduction in their size is desirable. Almost all of the routines are used to estimate the properties of pure compounds and mixtures and bear little resemblance to the needs of minerals extraction process modelling; similarly those compounds whose properties are stored in the ASPEN PLUS data banks are suitable only for chemical processing, the only one of interest to minerals engineers being water. If a mass balance only capability is required, this sufficing for most mineral applications, then it would be desirable to remove the entire property estimation section of the system; if however, an energy balance option is wanted then it would be useful to retain the basic structure and some of the utility routines, although it would be possible to remove the large majority of routines and data bank entries. Other than these removals it is likely that the other necessary changes would entail editing the System

Definition File and possibly altering those general purpose unit operation modules, such as the mixers and component separators, which it is decided to retain (465).

The ASPEN PLUS input language (see Appendix 1) is particularly flexible and easily used, and it is therefore expected that any alterations will be largely cosmetic and are likely to involve interfacing with graphical input or pre-written input files, rather than modifications to the language and its processor. However the output report file produced after a simulation appear to be less suited to minerals engineers. Whilst the user has a significant control over which sections of the report are written, the most important sections are those describing the performance of the unit operations and the stream compositions. Although the contents of the former are determined when the modules are written and can be easily altered, the stream report is written within the system and requires major improvements, especially in the description of solid flows; the parameters describing a mineral distribution are presently written apart from its flowrates and the format of simply listing stream compositions in a section away from the unit operations could be improved. Also information such as cumulative percentage finer or coarser than size limits and average mineral content is appropriate in the analysis of solids streams. A useful feature for minerals applications which is presently lacking, would be more appropriate analysis of the overall plant performance; at the moment only a mass and energy balance is written, which is not very useful for a minerals separation plant in which the mass of all components remains the same. A measure of the recovery of components to particular streams, and their grade or concentration, would be more useful, although also more difficult to calculate; one method of calculating this analysis would be to write a pseudo-unit operation module which received plant input and output streams as input.

Another possible addition to the present ASPEN PLUS system would be more substream attribute types and the definition of useful stream and substream classes containing the new attributes. The changes necessary include new tables to be added to the System Definition File and the modification of unit operation modules so that they can handle the new attribute types. Attribute types which could be added are a particle size - mineral fraction - surface property distribution for flotation modelling, a particle size - density - ash content distribution for coal preparation plant modelling and an attribute which could

describe the distribution of both grade and flow rate with particle size; it is also worth considering the desirability of a general-purpose attribute which describes the variation of particle size, mineral content and an arbitrary third property such as magnetic susceptibility, a surface property or radiometric content, although this would be more difficult to handle.

A unit operation-type module which will prove useful for describing the change in particle size through a plant, would alter the size ranges considered in a stream by changing the substream attribute type. The need for such a module is evident when one realizes how different the size distribution used to describe run-of-mine ore, is to that used to describe finely ground flotation plant streams. The major problems in developing such a module arise from the choice of a suitable continuous function to fit the discretized distribution of the input stream particle sizes, and regression of the discrete data to determine suitable parameters. Although functions such as Gates-Gaudin-Schumann and Rosin-Rammler are commonly used within the industry, no function has been found to consistently describe the entire size range of various distributions.

7.2.2 Cost Estimation and Economic Evaluation

Although the estimation and evaluation of mineral process costs and economics have been largely ignored in the technical literature, it would appear that the factorial and discounted cash flow techniques used in the ASPEN PLUS CES Subsystem are similar to those used in the industry and that few alterations would need to be made. Of these the inclusion of more appropriate inflation indices and default factor values would appear to need the most immediate attention. Also worthy of attention is the development of a module for the estimation of product prices when these are highly dependent on the product composition; at the present time the user fixes a price which is then unaltered by the performance of the process and the product composition, a method which is unsuitable for determining the revenue accruing to a mine from the sale of base metal concentrates, this being determined to a very large extent by the quantities of various elements present. Assuming that care is taken in its use, then with these additions the ASPEN PLUS costing and evaluation subsystem should provide a useful means of estimating process economics without any further alterations other

than the addition of unit operation cost blocks.

The major barrier to the development of suitable cost models is the lack of appropriate cost data; whilst providing a useful core of information the CIMM publication (409) is lacking in some important areas and requires some augmenting. It also is restricted to Canadian prices and so it would be desirable to develop cost data banks for other major mineral producing areas. Even if the basic cost models can be developed, the almost complete lack of published values for the factors to be used to determine project capital requirements will seriously hinder their application. There is also a similar lack of available information against which the use of the models and subsystem can be compared to test their reliability. Therefore a major effort is necessary to collect appropriate costing information and test the accuracy of the techniques used.

7.2.3 Operation Modelling

Over recent years a great deal of work has been carried out to refine the understanding of the physico-chemical phenomena underlying minerals extraction processes and to develop mathematical models to describe their operation; sufficient general knowledge is now available for many processes, to be confident that progress will not be significantly impeded by a lack of basic process knowledge and that mathematical models, albeit largely empirical, may be developed for most. Whilst it is desirable to use phenomenological models which can give some insight into workings of the process, many of those developed to date require parameter estimation from full scale testwork and so empirical models are necessary in some instances; another important factor in the choice of models to be implemented is that they should have been accepted, or are likely to be accepted soon, as feasible and reliable design tools. Thus in the choice of a crusher model it was decided to use a method based on Bond's law and manufacturers' literature, rather than a phenomenological model which needed to be adjusted according to the results of extensive plant testwork.

As was stated earlier there have been very few evaluations of the use of models such as those used, for the design of mineral extraction plants, and progress in the testing and application of modelling will continue to be slow due to measurement problems alone (14). One of the difficulties arising from this situation is that it is also hard to compare the relative performance of different models, both in comparison

with each other and when used in conjunction with each other. Therefore one of the major objectives of a test programme should be study of the absolute and relative performances of various unit operation models, not only in describing the action of a particular unit operation but also as one part of a simulation of the complex process.

Implemented Unit Operation Models

Unit operation modules were implemented on ASPEN PLUS to simulate crushers, ball mills, vibrating screens, hydrocyclones, flotation cells, agitated leach tanks, solvent extraction circuits and copper electro-winning tankhouses; this was in addition to the existing modules simulating solid-liquid separations, mixers, splitters and component separators. These have all been shown to be robust and useful in simulating a variety of processing operations except for the solvent extraction module which is unstable unless used under a fairly narrow set of stream conditions.

When considering the use of these, and any additional modules in a design and economic evaluation application, the major drawback at the present is the lack of equipment sizing and costing relationships such that the mass balance calculations may be interfaced with the financial calculations; as was suggested in the previous section, this should be one of the prime objectives of any further development of the methodology. It is likely that the absence of suitable equipment cost relationships will prove more of a hindrance to development than sizing algorithms, especially as most models rely on equipment sizes being defined before the simulation can occur. One exception to this is in the sizing and costing of material handling equipment such as pumps, conveyors and storage facilities; these are assumed not to affect the mass balance calculations, although implicitly, a further assumption is made that adequate materials handling capacity is available. The inclusion of these facilities in a process evaluation could be handled by developing separate modules which could access the simulation results and carry out the necessary sizing calculations, including the provision of such facilities in the relevant unit operation modules or by allowing for the extra costs through the addition of more factors in the capital and operating cost estimation algorithms. Another decision which must be made before cost modules are implemented is on the most appropriate method of calculating the consumption of media and reagents. A common

method is to estimate the rate of consumption per unit feed flowrate; however this rate often relates to circuit feed flowrate rather than the feed to an individual unit operation block. Suitable alternative methods may include treating such consumption on a plant section level rather than on a unit operation level or to define the consumption in terms of time rather than flowrate.

Whilst the models implemented to date are of use in simulating a particular circuit, it is still necessary for the engineer using them to manually manipulate variables to reach the required design conditions. Although ASPEN PLUS can be used to a limited extent in a design mode to automatically adjust model parameters to approach control limits on stream variables, it would be a very useful additional feature if some unit operation modules could be used to design equipment rather than just to simulate it. Examples of this would be for screen and hydrocyclone parameters to be adjusted to achieve defined cut sizes, and for the crusher and ball mill modules to design equipment to reduce material to a particular product size. These could be achieved within the existing modules and using the same model equations except that alternative input formats and calculation procedures would need to be employed.

The new crusher simulation module implemented is based on existing design methods and, other than the above deficiencies, should be suitable for use in its present form. The most likely area for improvement in the model is in increased experience and knowledge of the concept of crusher power rates and power efficiency; as was seen in section 6.3, the model is highly sensitive to the power input from the crusher and, especially as this is the least reliable of the empirical relationships, great care should be taken in its definition and verification.

The ball mill module introduced suffers from a similar inexperience in the use of the underlying model. The most critical areas where further understanding is required are the scale-up of laboratory data to industrial mills and the characterization of residence time distributions for such mills, though a possible further area is the measurement and definition of the breakage rate and breakage distribution functions; improvements in the first two areas are only likely if milling equipment manufacturers co-operate and contribute their accumulated experience and data. In the possible event of this phenomenological form of model proving unsatisfactory a new module using Bond's law design methods

could be developed and used as an alternative, though due to their empiricism and other limitations it is felt that this would be a rather retrograde step. However in the absence of suitable detailed models of rod mills, at the present similar methods would have to be used and so the inclusion of a module based around empirical scale-up procedures is likely however useful the present ball mill model may prove.

Whilst it may be assumed that the comminution models provide reasonable predictions of energy consumption and product particle size distribution, they assume that breakage rates are identical for each mineral composition and that the mineral distribution within a particular size class is the same after comminution as before. This assumption is obviously inappropriate and it shall be necessary to introduce a module which will provide a more reasonable calculation of liberation characteristics. It is suggested that this would be separate from the comminution models and would be similar in structure to unit operation modules; in this way it would be possible to define liberation characteristics either at a size reduction machine discharge or at the outlet stream of a size reduction circuit.

Due to the lack of liberation models it would be necessary for the module user to define the distribution of particles in mineral classes for each size interval or the distribution of particle sizes in each mineral class. However both of these assume a prior knowledge of the overall particle size distribution and so it is necessary for a method of adjusting the defined size distribution according to the actual distribution to be included in the module; for this method to have any value or reliability it would be necessary for the defined mineral distributions to be measured near the actual particle size distribution.

The major drawbacks of the hydrocyclone and vibrating screen models are their dependence on empirical relationships. However whilst it is likely that these relationships may be more reliably refined and calibrated using further testwork it is unlikely that these will be major and/or significant. The most suitable development approach is for more theoretically based models, such as Ferrara's (124) and Rose's (128) screen models, to be developed further so that they may be used in design. As was discussed in section 6.3 the hydrocyclone model could be further improved by reverting to Plitt's

original model formulation (138) so that assumptions need not be made about cyclone geometry and the effects of feed flowrate may be included.

It is generally accepted that flotation processes may best be modelled using an analogy to first-order rate kinetics. The major deficiency of this method lies in the accurate measurement of flotation rate constants for each particle size - mineral fraction class using laboratory testwork and whether or not this constant reliably represents the flow of material to the concentrate stream rather than to the froth above the pulp; no successful model describing the froth effects has yet been developed and so the onus is on the measurement of meaningful rate constants. Whilst the flow of solids from a cell is reasonably described using this analogy, the flow of water is calculated using even more empirical methods; the method used, which is to control the pulp density of the tailings from each cell, is reliable and robust but relies on engineering judgement and experience for its successful use. A factor which is relatively ignored in flotation modelling is that of the pulp density and general pulp environment; the assumption is made that the rate constants measured accurately represent the flotation process and that changing pulp conditions do not have a significant effect, or that if they do then the rate constants must be remeasured. Therefore the model allows rate constants to be measured for each bank of flotation cells, inherently assuming that these were measured under appropriate test conditions. The flotation bank residence time distribution is modelled as a series of perfectly mixed tanks, though it should be noted that these tanks need not represent individual cells in sizing and costing calculations as the ASPEN PLUS system allows the mass balance block values to be overidden in the cost block input data.

The module included to simulate agitated leach operations is based on second order empirical relationships whose parameters must be measured by the module user. In this way generality has been maintained and the model is not dependent on a particular leaching process model; if it is wished to simulate a leaching reaction using a particular phenomenological model, this model can be used externally to predict the parameters of the module equations. The major restrictions with the existing module are that pressure is not included as a variable and that the extraction of only one metal is modelled; however the inclusion of either of these modifications would only require slight adjustments to the module coding.

The module implemented on ASPEN PLUS to simulate solvent extraction mixer-settlers is generally inadequate and ought to be replaced. The original model was formulated according to the common practice whereby only the basic unit operations are modelled and that combinations of these are used to simulate any process configuration utilizing this flexible structure; however more recent practice has tended towards developing modules containing certain sub-processes, such as counter current decantation and solvent extraction, in order to simplify data input and enhance solution of the overall process model (466). For solvent extraction simulation the most suitable arrangement would be for a module to simulate each of the extraction, scrubbing and stripping sections separately, though the same module could be used in each case but with different operating parameters. Whilst the module may be based on similar calculation procedures as that existing, it is likely that a more efficient solution method may be developed utilizing the number of simular simultaneous equations present in the model. As well as a more efficient solution procedure further improvements needed in solvent extraction modelling are the development of extraction kinetics relationships, more theoretically based equilibrium models and equilibrium relationships for the scrubbing and stripping stages.

Additional Unit Operation Models

Extra unit operation modules which could be added to the present library with very little development work and which would considerably extend the present system capability include a general minerals separation module and a static leaching module, in addition to the liberation model described above.

The general module for the physical separation of minerals could be designed so as to simulate a variety of unit operations such as magnetic, gravity and electrostatic separators which cannot otherwise be modelled using less empirical techniques. As was suggested in Section 2.8 these separators may be best simulated using user-defined partition curves of the separation efficiency of the machine for a certain mineral property. A curve may be defined for the entire stream or for each size class by empirical functions of commonly used forms or, alternatively, by defining the separation efficiency of each particle class.

As was discussed in Section 2.5 the phenomenological models

proposed to describe static leaching of copper sulphide and oxide ores are unsuitable in form for steady-state process simulation and are generally not applicable to a variety of ores without alterations. It would therefore appear that for modelling a static leaching operation as a part of an overall process simulation a more flexible, empirical model would be suitable. The actual metal extraction could be modelled using a relationship similar to those employed in the agitated leaching module, the major variables to be included to be determined from use of the more detailed phenomenological models and from operating experience. In practical use the parameters for this relationship would be determined either directly from laboratory testwork or from fitting a model to data and then using the model in a planned series of experiments to determine the parameters. Whilst this is primarily designed to be of use in dump and heap leaching, it is likely that if a number of such modules were arranged together in a suitable fashion, it would also be possible to simulate a vat leaching process.

Extension to other Mineral Systems

The original selection of unit operations and stream structures to be implemented was made with the intention of modelling copper extraction processes, although their application in the modelling of other mineral systems was always considered. The only instances in which complete flexibility was inappropriate were those in which chemical reactions or transformations took place, and even then, when it was decided to use an empirical leaching model, it was practicable to develop this in a completely flexible fashion for use in modelling other leaching reactions. If the suggested additional unit operation modules and stream structures are implemented, ASPEN PLUS would be capable of handling all minerals separation processes, and if the general purpose modules already existing are used in conjunction with LEACH, then most hydrometallurgical extraction routes could also be modelled. Therefore, following the inclusion of these extra modules, the major development area should be the more detailed modelling of hydrometallurgical processes, especially when it considered that the analysis of such processes is one of the major advantages that ASPEN PLUS is likely to hold over potential competitors in the minerals extraction industry.

It would appear from the literature that very little theoretically-based modelling of solvent extraction processes has been carried out

with the application to design in mind; many of the relationships used in the modelling of solvent extraction for design have been of an empirical form and could thus be used in the same module as those used to model copper - LIX systems. The development of suitable relationships, if not available in the literature, could be based on laboratory determined data or could utilize data banks accumulated from various sources, such as that existing at Warren Spring Laboratory (329).

Although mathematical models of the processes involved in continuous ion exchange have been developed (467-470), there is still the need for testwork to determine kinetic and equilibria values for a particular system; whilst the similarity of these processes to solvent extraction should encourage the development of relationships which can be immediately used for design, it is still preferred to use empirical correlations obtained directly from the solutions under consideration (470). However the mathematical algorithms used in the models are based on established equipment design and scale-up procedures, and the parameters may be determined easily using small-scale apparatus (470-473).

Perhaps the most system specific unit of a hydrometallurgical process is the precipitation stage. Although more detailed models can be developed allowing for equipment parameters and precipitation reaction variables, it is also possible to model the operation using a general component flow separator. If it is assumed that for typical aqueous streams a precipitation technique will have a definable efficiency, then the only major additions to the ASPEN PLUS system will be cost modules to evaluate the various precipitation routes and possibly simple modules to allow for equipment parameters such as that implemented to model copper electrowinning; suitable processes to be modelled include copper cementation, yellowcake precipitation and the carbon-in-pulp method for gold extraction.

One advantage of the ASPEN PLUS stream structure flexibility which has not been utilized in this project, is its ability to model solid-gas and solid-gas-liquid streams. This potentially allows processes such as fluidized bed roasting and drying, sintering, pelletizing and cement kilning to be modelled and included in a process simulation. For such models to be implemented, it is likely that a major development effort will be required as an initial literature survey produced few references of note and it would also be necessary to write and use extensive physical and thermodynamic property estimation routines.

Before any major developments are made in introducing models of hydrometallurgical processes, it is important that detailed consideration is given to how energy balance calculations are to be made; the models so far introduced calculate only the mass balance around a unit operation. Whilst routines similar to those used in chemical engineering may be used to model aqueous streams, though it should be noted that presently, water is the only compound of interest whose properties may be estimated, the problem of how to define the heat capacity or enthalpy of a mineral mixture is likely to be difficult to overcome. Even if energy balance calculations are not carried out, the effect of temperature on unit operation performance should be included in each model.

The development of a module for coal washing operations could make use of the extensive efforts expended in developing process simulation in the coal industry; this is the one area of minerals extraction in which these techniques have flourished, partially because simple computational schemes are practicable due to the use of a simple, fairly rigid data structure and the relative ease with which the separation units can be described, and partially because of the advantages to be obtained when a large number of process flowsheets can be used to treat a coal, especially when a number of possible product mixes of different value exist.

The design of coal preparation plants has always been based on the laboratory determination of the specific gravity and size distributions of a plant feed and empirical estimation of the performance characteristics of washing devices. Pilot plant or even small-scale testwork is very rarely used as these techniques have proved to be very reliable.

The methods of describing washing machine behaviour are generally graphically based, using Tromp distribution curves, M curves or similar techniques; these have been described by Peng et al (474) and are similar to those proposed for the general separation model.

In recent years models of the actual processes have become common and have been used in performance prediction, project planning and process optimization; these models have also been reviewed by Peng et al (474) and Walters and Ramani (475).

There have been four main approaches to the modelling of coal separation processes, all of which involve a mathematical description of the distribution curve. These are:-

1. Representing the curve as a mathematical equation
2. Converting the curve to a straight line
3. Using a curve fitting routine
4. Using the probable error to describe the linear portion of the curve and an interpolation technique for the non-linear portion.

Due to the nature and shape of the curve, methods 2.- 4. are difficult to correlate accurately without the use of many parameters and quasi-constants. However a method which has proved applicable and meaningful is the generalization or normalization (476) of distribution data to characterize the performance of washers. This has been used as the basis for a flexible simulation program (129) developed in Pittsburgh and sponsored by the EPA and USBM; models have been developed of crushing, screening and flotation processes as well as of washing devices. The latter were derived from a series of reports (382) by the USBM publishing data on the performance of a number of cleaning operations throughout America; a similar later report would allow the addition of the BATAC jig to the program's capabilities.

A similar program has been developed in Canada based on accumulated experience in coal plant design (383). This includes representation of most coal washing operations, using extrapolation and interpolation routines to produce a continuous mathematical representation. Another similar program has recently been described by Laurila et al (384), although in this case mathematical equations have been derived using regression techniques to describe the washer separation curve; the use of equations rather than tabular data has allowed this program to be implemented on a micro computer, unlike the two previously mentioned programs which require more powerful computers. However it is only the USBM/EPA program which is publicly available, the others being of commercial origins and therefore used only in-house.

Another general purpose coal preparation plant simulator, SIMPREP, was developed by the National Coal Board and also contained individual models of unit separation processes (477); however this was regarded as "clumsy" and contained no economic features, and was therefore superceded by COALWASH (478). This latter program included product price values but only modelled float-sink devices in a considerably more general fashion and is more suited to initial design work rather than Type C project studies.

Whilst a different module for each coal washing device could be developed, it would be more efficient to develop a module common to all devices due to the similarity of the interfacing and reporting functions which would be required for use on ASPEN PLUS; it is felt that whilst these similarities suggest the development of a common coal washing module, similar to that used for the various crushing devices in CRUSH2, this is still a compromise and that it would not be advisable to include these models in the even more general purpose physical separation module proposed. Before implementation of such a module is considered, it would be advisable to consult potential industrial users on the most suitable stream structure; whilst a particle size - density distribution has been suggested, it is possible that ash or sulphur content may be desirable as additional stream variables. It should also be noted that although such a module would be a useful extra capability to more diversified companies and educational users, even a completely stripped down version of ASPEN PLUS is unlikely to be competitive with a program specifically designed for coal washing plant simulation.

7.3 Other Applications of the Methodology and Computer Systems

The proposed methodology is expected to be of more general application than just the Type C evaluation level and so it is worth considering the ways in which it may be implemented and whether the ASPEN PLUS system would still be of use.

For studies at earlier stages of process development, it would initially appear that the use of a system such as ASPEN PLUS would be limited or, rather, regarded as excessive. If however, as is more appropriate, the project development procedure is regarded as a continuous and dynamic process,the distinctions between the use of simple, linear process models and more detailed phenomenological models, and between order-of-magnitude and factorial economic evaluations, become less clear. The use of simple process models for operations such as solid-liquid separation can be combined, in the same simulation, as more detailed models, such as those used for leaching and solvent extraction; also, the general purpose ASPEN PLUS modules for component separators, reactors and mixers can be combined to simulate a complex flowsheet using predictions of unit operation behaviour and the material balances so derived, used in a factorial economic evaluation. The flexibility of the ASPEN PLUS costing and economic evaluation subsystem, especially the "hardwiring" of default values for cost factors, allows the calculation of evaluations from very little input other than the approximate equipment sizes. Whilst so detailed estimates would not be produced if manual techniques were to be used, the provision of such a facility, with such little necessary effort required for its use, enables these potentially more reliable evaluations to be calculated rather than depending on more generalized short-cut procedures; it should also be noted that the ASPEN PLUS economic subsystem does allow the input of plant cost correlations if a user still perfers to rely on such techniques. It would therefore appear appropriate to also use the ASPEN PLUS system for the less detailed process analyses, although its size and complexity would seem excessive if only this type of analysis was to be carried out and/or if the available computer hardware is inadequate. For these latter situations it could prove more desirable and efficient to develop considerably smaller systems capable of carrying out evaluations based on simple process models and cost correlations. Whilst programs have been developed for the material and energy balance

calculations (17, 18) these have not been combined with the cost and economic evaluation aspects of a process analysis; even if it is felt that such programs are more desirable than the use of a system such as ASPEN PLUS, then their development and use could be considerably eased and enhanced, utilizing the experience gained from using the general-purpose unit operation blocks and evaluation procedures on the ASPEN PLUS system. If however ASPEN PLUS is available on the type of micro-computer network environment envisaged in Section 7.2, then it should be more appropriate to use ASPEN PLUS for all evaluations, introducing more detailed models as more information becomes avaiable, until a feasibility study evaluation using phenomenological models and detailed factorial estimation techniques can be developed.

As process development continues after the feasibility study stage, through possible pilot-plant testing and optimization of the operating, full-scale plant, more detailed information will become available for analysis purposes and more accurate performance predictions will be required. The proposed evaluation methodology is just as valid during these development stages as before, although different process models would be needed to utilize the more detailed data and meet the more stringent accuracy requirements, and certain differences in the calculation techniques might be thought to be desirable. With respect to the latter, perhaps the most likely change would be the use of specially designed simulation programs, tailored for the process plant under study; whilst it is true that these could be more efficient for the simulation of a particular plant, suffering none of the overheads associated with a general purpose simulation system, they would suffer from the major disadvantage of being unable to analyse the effects of changes in the process configuration. Another possible alteration to the calculation procedure could include changes in the ASPEN PLUS economic evaluation subsystem; the capital cost estimation routines could, however, easily handle the availability of more detailed cost information through the input of actual equipment and plant costs and the over-riding of the default factorial estimation procedures. The most likely change to the economic calculations would be the inclusion of procedures for evaluating the profitability of process configurations, when the plant is already constructed and has been operating for some time; depending on the detail required this could prove to be as simple as removing the need for capital cost estimations to be calculated.

Almost certainly the major changes to the present ASPEN PLUS system for its use in more detailed process analyses would be the addition of more unit operation models. As the models presently implemented were designed to be of use in the feasibility study stages of process development, only those that utilized data which would be available at that development stage were considered. As the amount of available information increases the use of models which considered more aspects of the process phenomena would become desirable. For instance the use of the present crusher model could be superceded by the implementation of a more detailed model including variables such as crusher gape and internal classification; the type of model would probably be similar to Whiten's (92), including modifications similar to those used by Hatch and Mular (94). Other models to be added may include a screen model, such as Ferrara (124). Lynch et al's hydrocyclone model (236) and a flotation model which included more details of mixing and froth effects. The actual models implemented will vary from installation to installation, depending on the particular plant performance, and also as more reliable model forms are developed.

Other additions which might prove useful would be associated with the collection of process data and the transformation of this to provide ASPEN PLUS model parameters. In the testing and calibration of unit operation models it is essential to have data with good mass conservation properties; however in the analysis of minerals extraction processes, a great deal of the collected data are often redundant and erroneous due to natural disturbances, sampling errors, unreliable instrument readouts and laboratory analyses inaccuracies. Therefore various complex methods of data processing (479-482) have been developed in which most available data collected around a circuit can be used to compute an optimal material balance associated with a confidence interval for the calculated flow rates and assays. The various solution procedures used to calculate a consistent mass balance have recently been reviewed by Mular (483), whilst Reid et al (484) have surveyed the available computer software packages for their automation. The use of such a package would probably be very useful in the testing stages of the ASPEN PLUS system and models proposed in ch. 7.2, and the interfacing of the two packages could prove a useful feature when operating plants are to be simulated on a regular basis. However such a material balance program would not directly provide model parameters and so part of the

interfacing with the simulation system would be likely to include data regression routines to calculate the model parameters from the material balance data; these routines would be of similar form to the parameter estimation routines developed for the NIM flotation model (247) and would need to be derived for each unit operation model.

When considering the development of a version of the ASPEN PLUS system for the minerals industry, it is perhaps worth considering that as a general purpose process engineering system, its application is not restricted to minerals extraction processes; the system's capabilities in steady state simulation could well be utilized in the evaluation of mine ventilation and refrigeration networks, and of the supply of utilities such as power, water, steam and compressed air to the various locations in a mining complex. As with those system applications previously discussed, each of these problems could potentially be analysed using programs specifically designed for the purpose, but in the absence and/or development of such programs the use of a flexible system could prove of value.

Yet another application for a flexible simulation and economic evaluation package such as those proposed, is in the development of simple empirical relationships describing important process variables; these could be of a similar, though possibly more detailed, form to those already published by O'Hara (412) and the USBM (411), and may include the effects of feed tonnage and grade on process recoveries, revenues and costs. If a reliable model of a process is developed using computer simulation techniques, that model could be used, in an analagous fashion to laboratory equipment, to perform experiments to study the effects of major variables; if these experiments were carried out in a suitable fashion, it would then be possible to develop a regression model describing these major variable effects. The development of a number of such models would be greatly enchanced using computational methods and would be of considerable value within the minerals industry.

7.4 Future Methodology and Systems Development

In the previous two sections, specific improvements and development areas have been suggested to extend the application capabilities of the methodology and its supporting computer system, and to improve reliability of the mathematical models used. Almost certainly, the most important facet of any such development programme would be the industrial testing of the methodology, especially the models used, and the ascertaining of statistical confidence limits for the simulation output; it should be noted that the results of this testing should not be regarded as final and complete, but, rather, that such testing of the accuracy of the simulation should be progressive so that confidence in the technique will improve as it is used and refined further. Especially if a single system such as ASPEN PLUS is used to implement the various levels of process study, once sufficient experience has been obtained in the reliability of the models, it would be advisable to include in the program output some indication of the probable accuracy of the particular models used for a simulation.

The suggested improvements and developments to the methodology and computer system should only be regarded as possible specific changes, and a part of a more general approach to methodology development and implementation. Now that it has been shown that tools and techniques necessary to utilize the concept of the methodology are available, it is necessary to consider how its widespread and responsible use may be promoted and developed.

Progress towards accepting modelling and simulation techniques in the minerals extraction industry has been very slow; however this slow acceptance has been a cause rather than a consequence of the limited success of the techniques' various applications to date. The major cause of the difficulties in their implementation is rooted in the multidisciplinary nature of modelling and simulation which does not suit the traditionally conservative approach to plant design that prevails in the industry; the adoption of methods involving a wide scope of disciplines has met some resistance in minerals extraction, a field where operations have sustained a relatively low rate of innovation and suffered from a conservative attitude to change (479).

For successful implementation of the methodology it will be necessary to match the computational techniques and models used with the practical problems of obtaining and utilizing all accumulated data

and knowledge at all stages in the process development programme. There must be real and full collaboration between the engineers, whose everyday work is the design of real processes, and those whose task is to develop the computer techniques and computer design aids which the use of the methodology offers; if there is not this collaboration then the full potential of the methodology and its associated computer systems will not be fully realized.

Evans (428) has outlined five distinct stages in the life cycle of any large scale computer package: definition, design, implementation, testing and delivery; for large flowsheeting systems such as ASPEN PLUS, the time elapsed between definition and delivery is about seven years. Obviously, through the acquisition of existing software such as ASPEN PLUS, the development time for a system as proposed in this thesis will be considerably reduced; however even with the present state of development and using ASPEN PLUS, then this period is likely to be of the order of 3 - 4 years assuming at least 2 or 3 full-time workers engaged in the project.

In the project definition stage it is essential to determine what is to be included in the proposed system, what type of analyses it is to perform and what characteristics of user interface are important. During this critical stage it is therefore important to establish a close relationship with the ultimate users of the system and therefore determine what it is that they want. However, considerable judgement is necessary in this task as these users may not recognize what is realistically achievable and will therefore tend to be unduly conservative, asking only for what is familiar; they will resist changes from their accustomed ways of studying a problem and from using the techniques and knowledge available to them (428). In this respect the work carried out to date may help in reducing the barriers to the introduction of such "new" technology.

A major problem which is likely to arise during the project definition phase is an ignorance of the computing environment which is likely to exist in the late 1980's when the project's efforts should be coming to fruition. As has been suggested in Section 7.2, it is likely that most engineers will have access to a distributed computer system with considerable local power and the ability to call on larger data processing facilities almost at will; the engineer utilizing such a computer system would probably be using a desk top terminal/personal

computer with graphics capability.

In such an environment it would need to be decided how to distribute the modelling functions on a distributed network. For instance, at what point does the user switch from a small tailor made simulation routine, such as calculating the mass balance around a simple flotation circuit, to a larger general purpose flowsheeting package, which can handle regrind circuits, size classification and dewatering as well as evaluating the process economics; especially important is the manner in which those modelling functions carried out locally, may interface with the larger, centralised flowsheeting packages. As the use of process mass balance and economic modelling systems increases in the minerals extraction industry, so also should the use of integrated computer systems increase, with the output of one program serving as the input to another; an example of such an integrated system suggested (428) in the chemical engineering industry includes: the results of steady state simulation being used as the initial conditions for a dynamic simulator: the results of detailed process modelling being used to update simpler models (the concept of the meta model, which is likely to be especially useful in cost modelling): flowsheeting systems used as an integral part of a data regression capability to fit parameters in flowsheet models with plant data: the use of probabilistic techniques rather than deterministic, to provide estimates of economic risk or probable performance, and their statistical variance.

Whilst discussing the distribution of process modelling functions and the integration of process analysis techniques around a centralized computer package, it is also worthwhile considering how these may interface with the other phases of mineral project evaluation and development, such as exploration, mining, smelting, refining and the provision of supporting infrastructure. As was suggested in Section 5.2, in order to obtain a complete analysis of the economic factors affecting a processing plant, it would be necessary to include full details of all of these other phases and their cash flows; similarly it is necessary to consider the initial mineral processing. For instance, the relative effects of mining costs may be ignored in comparing a flotation plant and an agitated leach plant; if however, dump or heap leaching was to be studied, then it would be necessary to include the effects of changes to the mine design such as higher tonnages, lower grade and different

haulage routes. As an initial screening tool it may prove useful to include simple pseudo-unit operation blocks to simulate the mining process; these could be based on published correlations of mine equipment sizes and costs (e.g. 411). There is however a danger that these would be imbued with undeserved importance and accuracy, and/or that they would be incorrectly used by unqualified personnel. It should always be remembered therefore that the main advantages of the process evaluation methodology are in comparing alternative flowsheets and identifying the economically sensitive areas, rather than providing an absolute measure of a process' economic worth.

Whilst the development of a similar integrated system for the minerals extraction industry would appear revolutionary to many, the results of the work in ASPEN PLUS shown in this thesis and the developments proposed in the preceding discussion, indicate that most of the system components are already, or at least nearly already, available. Although the validity and utility of an integrated modelling approach to process design and analysis may have still to be proven, especially to more conservative elements, the use and integration of many of the components will be necessary during this testing and proving phase and as such, an integrated system should be the ultimate aim of any further development of the methodology. Even though an industrial organization could conceivably develop such a system, the skills and experience in the various disciplines required are unlikely to be concentrated in one organization and so it would be necessary to import the relevant knowledge and personnel. As it would also be advantageous in terms of compatibility and portability for potential users to concentrate on a common system, the preferred development procedure would be for one central establishment, probably academic, to co-ordinate the procedure with perhaps specialized sections, such as unit operation modelling or data balancing, to be developed by satellite bodies; in this respect it is worth noting that the modularity of such a system should allow an organization's unique procedures and expertise to be incorporated into their own version of the system without prejudice to their confidentiality or the basic compatability of the methodology and computer systems.

Another major issue to be discussed during the system definition phase of the development programme is the nature of the user interface,

this encompassing both the system input and output. As was described in Appendix 1, the ASPEN PLUS input language is very powerful and flexible, and may be considered as the state of the art. Questions that may be raised should include whether or not graphical input facilities are useful; whilst these could be utilized to input process topology and graphical correlations not restricted by mathematical form, there would still be a very large part of the input data which is of a numerical nature. As well as using the insert file capability of the ASPEN PLUS language to make process input even simpler, use could be made of the increasing availability of full screen editing by perhaps providing a fill-in-the-blanks form for the user to edit and alter the parameters on commonly used process flowsheets. The nature and composition of the system output also needs consideration; as was discussed in Section 7.2 the present ASPEN PLUS output reports are perhaps more suitable for chemical processes and the requirements of the mineral industry users will need to be determined. Also, whilst it is presently possible to control the amount of report output, there is no flexibility or choice in the report contents, a situation which should be improved.

Almost certainly the most important decision to be made during the system definition phase concerns the choice of the executive software. At the time of writing the ASPEN PLUS system is the only proven and reliable system capable of handling the range or processes likely to be encountered in minerals extraction, and as such it is assumed that it would be chosen. However it is important to note that the general trend of process flowsheeting software is towards the greater use of the equation-oriented approach (428). Initially it is envisaged that simultaneous modular systems will replace traditional sequential modular methods; modern systems such as ASPEN PLUS are, though, designed to be used using both approaches, simply regarding the simultaneous solution of all stream variables as another form of stream convergence. Towards the end of the decade it is likely that completely equation-oriented systems of the SPEED-UP type should be approaching commercial use (455) and will gradually supercede the previous ASPEN PLUS type of program; however as so much effort has been invested in the development of software for the present modular systems, the new programs will need to be able to absorb existing routines (455).

Although it is envisaged that potential industrial users would be involved in the project definition and design phases, one of the largest barriers to acceptance of the methodology is the conservatism of minerals companies towards new technology of this sort and so the problem of industrial transfer should be addressed as soon as possible (479). Whilst much of the software is likely to be written during the testing and validation phase, rather than beforehand, it would be advisable to start the process of educating users in the use of the methodology and the computer systems as soon as is practicable. It has been found in the chemical engineering industry that testing is one of the most costly and time-consuming stages in the life-cycle of system development, and that the only technique is to run numerous benchmark problems and persuade people to use the system as much as possible (428). There is evidence (479) in the field of mineral process modelling that some workers are already developing methods to overcome this problem, experience to date showing that whilst the heart of any computer software transfer is good, sound documentation, important tools also include industrial/academic workshops, tailor made packages and the use of university students (479). As has already been suggested industrial-academic interaction is an essential part of the system definition phase and should continue throughout, and after, the development programme. The use of tailor-made packages verifies the suggestion that the testing and validation phase, if carried out in conjunction with potential users, is likely to form an important part of the methodology transfer process. Whilst the use of the system to simulate and evaluate a company's process would be extremely useful, it is also feasible that from these simulations could be derived smaller, but less flexible programs for use in evaluating a particular process; although perhaps retrograde to the introduction of a general purpose system, this could encourage more use of the techniques proposed and lead people on to the use of larger, more flexible systems.

The almost immediate introduction of the use of computer-aided design and evaluation methods and systems to the education of students of minerals extraction processes is likely to be of value not only as an educational tool but also as a means of introducing the techniques to industry. As well as graduating with an understanding of the methodology and its potential benefits, students can be used extensively

during industrial training periods to implement the computer systems at industrial sites, experiment with them in a working environment and illustrate their use and value (479).

7.5 Summary and Conclusions

Existing manual calculation methods for the design and analysis of minerals extraction processes are inadequate to meet the more strenuous modern day demands of the present economic climate and world-wide diminishing resources. Therefore a methodology has been proposed for the computer-based evaluation of minerals extraction processes. This methodology uses models of the constituent unit operations to develop an overall process model, including details of the mass and energy balance,equipment sizes and costs and overall process cash flows. Due to the roles of practical experimentation and experienced engineering judgement in plant design, it is not intended that the methodology should replace existing design procedures, but rather that they be enhanced and that more comprehensive analysis may be allowed within the strictures of a development programme.

This methodology may be applied at different stages of process development, though each level of application would require varying amounts of included detail in the unit operation models to take account of the quantity of data available. Methods by which each level may be implemented have been described, but for the purposes of beginning the methodology development the feasibility study type of process evaluation has been concentrated on. The available design techniques, process models, costing and evaluation techniques, and executive computer systems have been studied, and the implementation of a suitable supporting modelling system initialized and illustrated.

From these studies it is apparent that the technology required to develop such a system is rapidly becoming more readily available, but is still not extensively used within the mineral industry. Even though the initial development of the system depends largely on proven design technology, a major objective of further development will need to be the testing and validation of the modelling approach and the transfer to and acceptance by industry of the methodology and supporting systems. Methods by which this transfer may take place and steps to be taken to smooth its way have been suggested.

The ASPEN PLUS flowsheeting system has been used for the development and support of the methodology. This is the state-of-the-art process simulation system, though it has required the work in this project to enable its application to minerals systems.

The work involved improvements to the stream data structures and the introduction of unit operation models for the simulation of crushing, grinding, screening, hydrocyclones, flotation, leaching and electrowinning processes. Using the resulting capabilities it is possible to calculate mass balances for a wide variety of minerals extraction flowsheets; because only restricted time was available to use ASPEN PLUS, the applications, and the limitations, of these models were illustrated using specially designed programs. The costing and economic aspects of the methodology were also demonstrated using a program developed for the purpose. Further changes which could be made to extend the present limits of application include the addition of new and improved unit operation mass balance and costing models, and peripheral alterations such as improved report writing and simplified input methods for the completely inexperienced users. Suggestions have been made as to where these changes may be most profitably made and also how the system may be further developed as a part of an integrated system for mineral project evaluation.

Appendix 1 The ASPEN PLUS System

The ASPEN PLUS system for process simulation and economic evaluation is a commercially maintained, debugged and updated version of the ASPEN (Advanced System for Process Engineering) system. The original ASPEN software was developed between 1976 and 1981 by workers at Massachussets Institute of Technology under contract to the U.S. Department of Energy. Whilst the immediate intention of the project was to provide a method for the technical and economic analysis of proposed synthetic fuel projects, it offered a unique opportunity to develop a next-generation simulator that would extend the technology of process simulation to cover a much broader range of processes than was previously possible (438,439). Funds were made available to support a full time staff of 15 and to acquire proven industrial software, and an advisory committee, containing representatives of 65 companies and government groups, was formed to determine the requirements of the practising engineer for process engineering software. Program development followed a structured approach for the first time in such a system and this, combined with the use of proven technology, produced a unique system with considerable capability and flexibility. After a two-year testing stage involving 55 industrial and government user groups, the ASPEN system was delivered to, and made available by the Department of Energy in the form of source code and documentation. However the testing has not removed all of the systems faults or problems and the publically available version will not be supported or maintained. Another version of ASPEN, called ASPEN PLUS, has been made available on a subscription basis from Aspen Technology Inc., a company formed by members of the M.I.T. team after the original project ended. This company offers maintenance and consultancy on this new version, periodically produces improved versions of the source code and supplies training in the use of ASPEN PLUS.

System Architecture

ASPEN PLUS is written in a sub-set of FORTRAN IV to ensure portability; to-date it has been installed on a variety of computers, the only major engineering computers unsuitable being Control Data Corporation machines. Whilst Fortran is inefficient compared to languages such as PASCAL and FORTRAN 77 for implementing list-type data structures and for processing character strings, it is available and maintained on nearly all computers and trained personnel to implement programs in FORTRAN are also commonly available.

The system's flexibility, and the source of its uniqueness, stems from its utilization of a plex data structure of the type proposed by Evans et al (22). In the plex, information is stored in blocks of contiguous locations known as beads. Beads of any length are created dynamically from a pool of free storage which is actually a lengthy FORTRAN array named PLEX, and may contain integer values, real values, pointers to other beads or character strings. During the initial processing stages of an ASPEN PLUS simulation run a Problem Data File (PDF) is written structured as a plex, and this contains all of the data required to carry out the simulation and storage locations in which to write the results. The first bead in the plex is known as the Process bead and this acts as a directory containing the locations of all beads in the file. Other than physical property parameters, which occupy a major section of storage space in a typical chemical process simulation, the bulk of the data is stored in either a stream bead or a unit operation bead, simple examples of which are shown in Figs. A1 and A2.

Beads are identified and stored by bead number so that their contents can be accessed by use of the bead number and system data management subroutines to determine their location in the PLEX array. As the plex structure is rather cumbersome and inconvenient for unit operation and

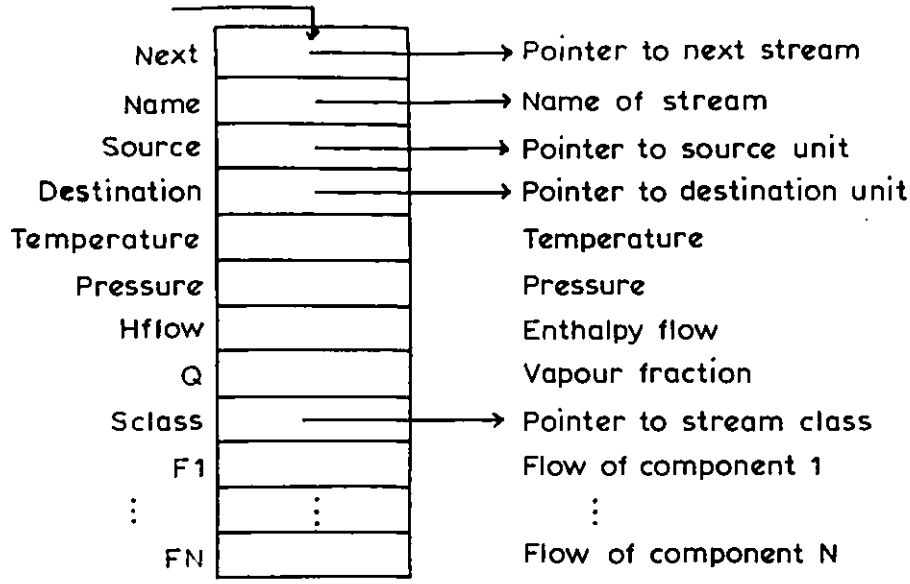


Figure A1 Bead to contain stream variables

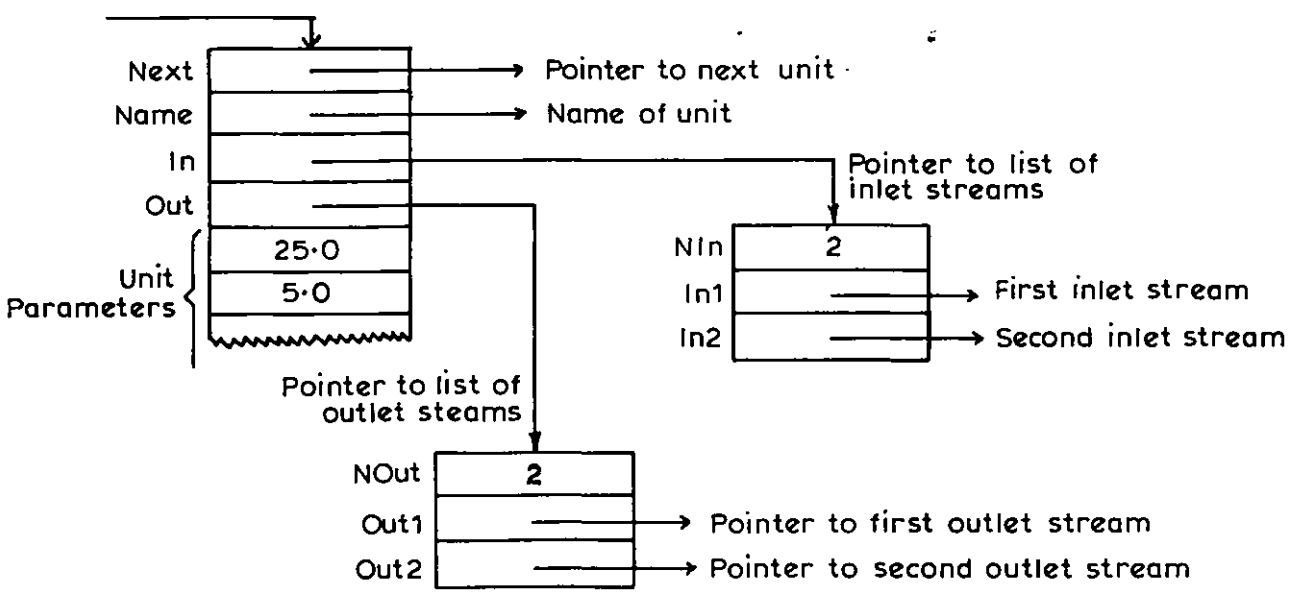


Figure A2 Bead to contain unit parameters

inconvenient for unit operation and other lower level routines, there are further data management subroutines for transforming data in the various classes of bead into arrays for easier manipulation and vice versa once the manipulations are complete; as will be seen the primary role of unit operation interface routines is to access and arrange the data from stream and block beads into arrays, and then to write the resulting data into the respective bead locations.

The potential advantages of the plex data structure have been utilized by adopting a preprocessor type of structure in which an input translator generates a main calling program which is then executed to perform a particular simulation. This input translator processes the user input, enters all of the data regarding the process into the Problem Data File and writes the main calling program containing the necessary calls to models; this is then linked together with the object modules from user and ASPEN PLUS program libraries and compiled. In this way a load module is created as a tailor-made simulation program for a particular problem. Because the input translator not only writes the calling program but also dimensions the plex and writes the process bead, there is a complete absence of any dimensional constraints on the system other than those imposed by the limitations of the computer hardware; thus there is no maximum number of streams, components, blocks, stages in a CCD circuit etc. The use of the plex structure, preprocessor structure and data management systems also allows complete flexibility in the higher level executive routines and dimensioned FORTRAN arrays in the lower level model routines.

Fig.A3 describes the flow of information in executing an ASPEN PLUS simulation and hence the role of certain major parts of the system. The Input Translator processes the user input, which is written onto a file in a relatively free format according to the rules of the special input language; having, literally, translated this input the input processor writes the main calling program, dimensions the plex, or problem data file,

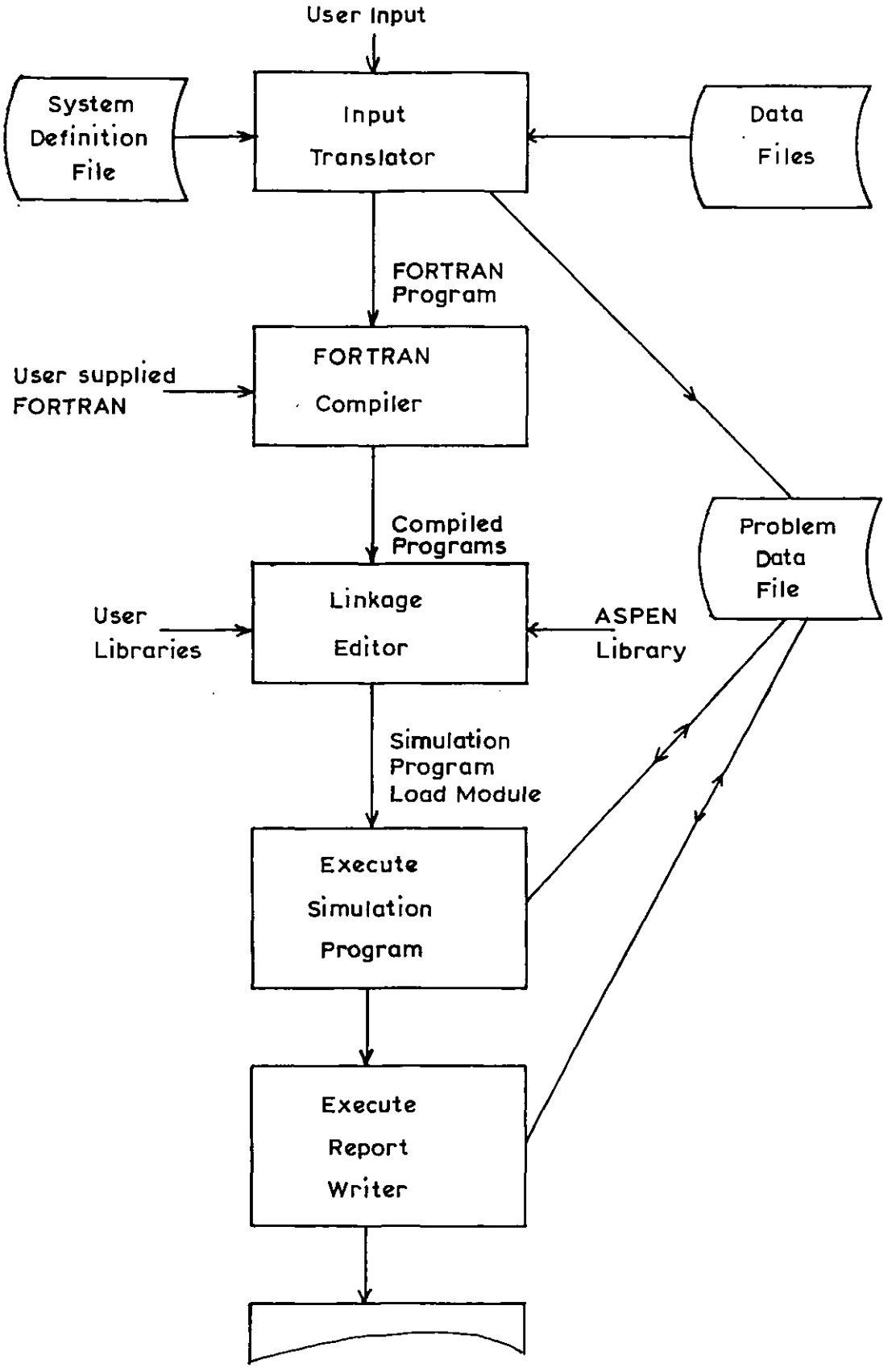


Figure A3 Flow of Information in ASPEN

and enters the process data into this file. The calling program is then compiled and linked together with the necessary object modules to form a load module. This module, or program, then reads data from the problem data file and after performing the necessary calculations writes the results back onto the same PDF; the report writer can then reproduce reports from the data in this file. Two output files are produced from each simulation run, a report file and a history file; the latter contains information describing the progress of the execution of the simulation whilst the report file is only written when the simulation is complete.

The Input Translator is table-driven, which means that all of the information required to process input statements (keywords, default values etc) is stored in tables in a file called the System Definition File (SDF). A table will be written to define each stream and substream attribute description, stream class definition, unit operation block etc. and so the type of process that the system can simulate is only restricted by the tables in the SDF, thus if it is necessary to add a new unit operation or substream attribute, the only system change which has to be made is the addition of a table to the SDF. By changing the contents of the SDF it is possible to change the applications of a particular ASPEN PLUS installation; thus if a mineral processing only version is required then it is a relatively simple matter to write a system definition file containing only the relevant unit operation and stream types.

The input language has been designed for process engineers who do not necessarily have an extensive knowledge of computer programming. The input can be considered to be made up of paragraphs, sentences and words; a paragraph begins with a primary keyword and may consist of one or more sentences. Each sentence begins with a secondary keyword that indicates the category of data appearing in the sentence; tertiary keywords are used to enter data and their values are the data items. For instance

from Fig.A4, in the statement:-

```
BLOCK B2  SCREEN2

     PARAMETERS SQ-MESH-APER  =  0.0095
```

the word BLOCK is a primary keyword indicating that the paragraph contains block data. The user-specified block identifier is B1 and the unit operation model to be used is the crusher model developed in this project, which is indicated by the use of the secondary keyword SCREEN2; model parameters, such as the number of crushers arranged in parallel, are defined in a PARAMETERS sentence, through the use of the appropriate tertiary keywords; the correct format for the definition of model parameters is described in the model documentation, as are the default values for optional paramaters.

Fig.A4a is an example of a complete input file for an ASPEN PLUS simulation run; the six essential paragraphs in any such file are the type of file (NEW/EDIT/INSERT), and the FLOWSHEET, STREAM, COMPONENTS and BLOCK paragraphs. Any other paragraph such as UNITS, REPORT, CONVERGENCE etc are optional and default values, as defined in the SDF and described in the User Manual, will be used if they are omitted. The FLOWSHEET paragraph describes the process configuration in terms of the constituent streams and unit operation blocks each of which must be assigned a unique identification label; the flowsheet described by this input file is shown in Fig. A4b. The individual compounds flowing through the process are defined in the COMPONENTS paragraph and their feed rates in the STREAM paragraph. As was suggested earlier the types of unit operation models to be used and their parameters are controlled by the use of BLOCK paragraphs. The use of the ASPEN PLUS system and the format of the input file and its constituent keywords are described in the User Manual; this manual also describes the use of insert files, these being set up for commonly required process flowsheets to make simulation runs even easier.

```
NEW
TITLE 'ALLIS-CHALMERS BENCHMARK PROBLEM'
IN-UNITS SI
OUT-UNITS SI
;
FLOWSHEET
  BLOCK B1 IN=S1             OUT=S2 S3
  BLOCK B2 IN=S2             OUT=S4 S5
  BLOCK B3 IN=S3  S5  S6     OUT=S7
  BLOCK B4 IN=S7             OUT=S8
  BLOCK B5 IN=S8  S16        OUT=S9
  BLOCK B6 IN=S9             OUT=S10 S11
  BLOCK B7 IN=S10            OUT=S12 S13
  BLOCK B8 IN=S11 S13 S14    OUT=S15
  BLOCK B9 IN=S15            OUT=S16
;
CONVERGENCE CONV1 WEGSTEIN
  TEAR S9
  PARAM MAXIT=60
;
COMPONENTS ORE /H20 H20
ATTR-COMPS ORE CAUSR1
STREAM S1
  SUBSTREAM MIXED MASS-FLOW=100.0 TEMP=300.0 PRES=1.0 &
                FLASH-OPTION=NOFLASH
                   MASS-FRAC H20 1DO
  SUBSTREAM NCPSD MASS-FLOW=1500.0                    &
              TEMP=300.0 PRES=1.0
  MASS-FLOW ORE 1500.0
  COMP-ATTR CAUSR1 3000.0
  SUBS-ATTR PSD FRAC=0.15 0.15 0.10 0.20 0.1 0.1 0.06 0.08 0.06
STREAM S6
  SUBSTREAM MIXED MASS-FLOW=100.0 TEMP=300.0 PRES=1.0 &
                FLASH-OPTION=NOFLASH
                   MASS-FRAC H20 1DO
  SUBSTREAM NCPSD MASS-FLOW=0.01                      &
              TEMP=300.0 PRES=1.0
               MASS-FRAC ORE 1DO
  COMP-ATTR CAUSR1 3000.0
  SUBS-ATTR PSD FRAC=0.15 0.15 0.10 0.20 0.1 0.1 0.06 0.08 0.06
STREAM S14
  SUBSTREAM MIXED MASS-FLOW=100.0 TEMP=300.0 PRES=1.0 &
                FLASH-OPTION=NOFLASH
                   MASS-FRAC H20 1DO
  SUBSTREAM NCPSD MASS-FLOW=0.01                      &
              TEMP=300.0 PRES=1.0
               MASS-FRAC ORE 1DO
  COMP-ATTR CAUSR1 3000.0
  SUBS-ATTR PSD FRAC=0.15 0.15 0.10 0.20 0.1 0.1 0.06 0.08 0.06
DEF-STREAMS MIXNCPSD ALL
DEF-SUBS-ATTR PSD PSD
 INTERVALS 9
 SIZE-LIMITS 0.080/0.070/0.060/0.045/0.030/0.020/0.012/0.008/0.004/0.0
```

```
BLOCK B1 SCREEN2
  PARAM SQ-MESH-APER=0.016 &
        DWIR=0.006 &
        SCREEN-AREA=118.0 &
        THETA=0.349 &
        SCREEN-TYPE=1.0 &
        BULK-DENSITY=1602.0
BLOCK B2 SCREEN2
  PARAM SQ-MESH-APER=0.0095 &
        DWIR=0.004 &
        SCREEN-AREA=118.0 &
        THETA=0.349 &
        SCREEN-TYPE=2.0 &
        BULK-DENSITY=1602.0
BLOCK B3 MIXER
BLOCK B4 CRUSH2
  PARAM  PINS=373.0 &
         TOND=194.45  &
         BWI=55.54    &
         MPOS=2       &
         CTYP=3       &
         MTYP=3       &
         CIRC=1       &
         NOC=7
BLOCK B5 MIXER
BLOCK B6 SCREEN2
  PARAM SQ-MESH-APER=0.016 &
        DWIR=0.006 &
        SCREEN-AREA=260.0 &
        THETA=0.349 &
        SCREEN-TYPE=1.0 &
        BULK-DENSITY=1602.0
BLOCK B7 SCREEN2
  PARAM SQ-MESH-APER=0.01150 &
        DWIR=0.004 &
        SCREEN-AREA=260.0 &
        THETA=0.349 &
        SCREEN-TYPE=2.0 &
        BULK-DENSITY=1602.0
BLOCK B8 MIXER
BLOCK B9 CRUSH2
  PARAM  PINS=373.0 &
         TOND=194.45  &
         BWI=55.54    &
         MPOS=3       &
         CTYP=3       &
         MTYP=3       &
         CIRC=2       &
         NOC=11
SIM-OPTIONS  ENERGY-BAL=0  SIZE-RESULTS=0
BLOCK-REPORT NOTOTBAL NOCMPBAL
RUN-CONTROL MAX-TIME=200
```

Figure A4(a) ASPEN PLUS Input File

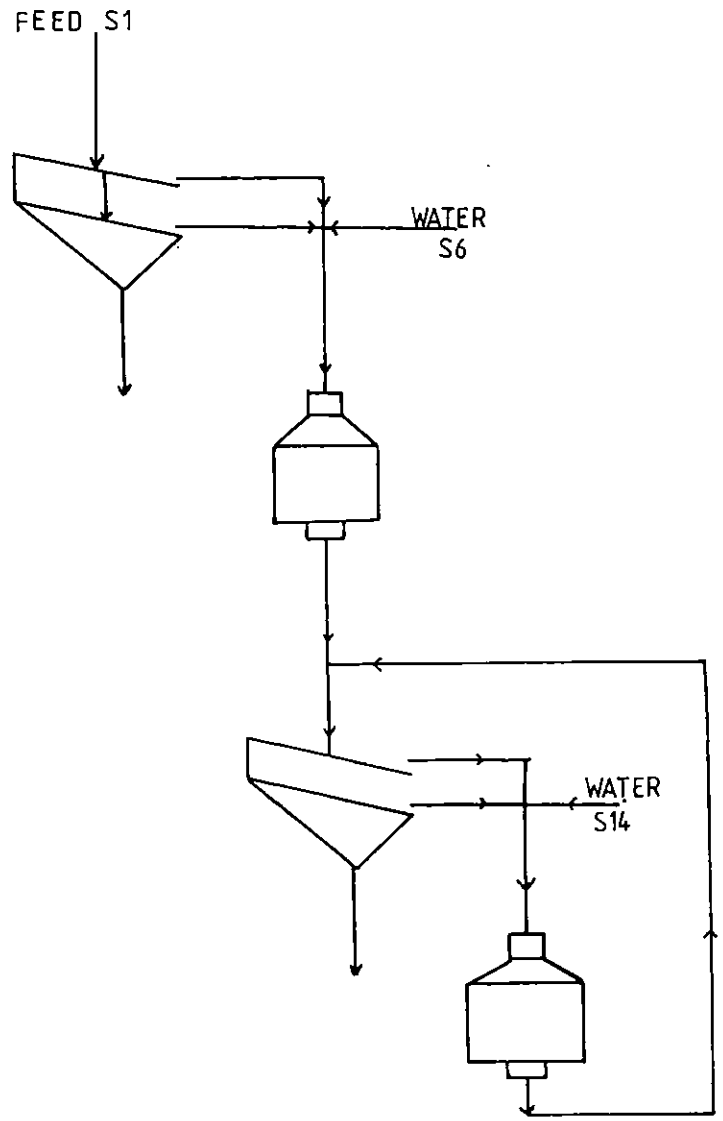


Figure A4(b) Crushing and screening circuit flowsheet described in Figure A4(a)

Stream Structure

The purpose of streams in any process simulator is to represent the flow of material between process units and into and out of the system, whilst the purpose of blocks is to describe the action of the process units on their input streams and to transform these accordingly to produce outlet streams; therefore the information that must be carried in a stream is that information needed by blocks in order to carry out their transformations on the inlet streams and so stream characterization is driven by block requirements. In general, the data required are the bulk system composition and two specifications such as pressure and temperature or pressure and enthalpy; for certain flow sheets other information such as particle size and composition is required, although the nature and detail of this information will vary between applications.

In the ASPEN PLUS system stream composition is described in terms of component flowrates, two types of components, termed "conventional" and "non-conventional" being used to characterize the full range of compounds from ore components to nebulous materials such as gangue or ash. Conventional components are typical of components found in traditional chemical engineering simulators and are used to represent pure compounds or pseudo-compounds that may be characterized in terms of standard pure properties such as molecular weight, critical pressure, vapour pressure coefficients, heat of formation and ideal gas heat capacity coefficients. The usual thermodynamic and transport properties of these components, their mixtures, or components in a mixture may be calculated using a variety of conventional methods. The efficiency and accuracy of a simulator describing the processing of conventional components generally depends largely on the physical property data banks and calculation subroutines, which also usually comprise a considerable proportion of the size of such a program; in the ASPEN PLUS system they consist of almost 50% of the program.

Nonconventional components are used to represent those materials which cannot be characterized in terms of standard pure compound properties. Instead they are characterized by vectors of data called "component attributes" which are state information required by nonconventional property models in order to calculate physical properties of the components or are direct definitions of those properties; component attributes can be analyses, such as ash or moisture content, or properties such as density. Each nonconventional component in a simulation may be characterized by its own set of component attribute types. The particular combination of component attribute types required, depends on the properties to be calculated for that component and the models used; for instance in the simulation of a coal gasification process coal would be described in terms of its ultimate (% carbon, hydrogen, nitrogen, sulphur, oxygen, chlorine and ash), proximate (% water, volatile matter, fixed carbon and ash) and forms of sulphur (pyritic, sulphate and organic) analyses, whilst for a flotation plant simulation a mineral or group of minerals might be characterized only in terms of its density. Whilst conventional physical propeties are stored in data banks and are held to be variable around a process, component attributes can change from stream to stream and are therefore carried as stream data.

ASPEN PLUS streams are subdivided into one or more substreams, each representing a portion of the stream that is to be treated in a special manner by unit operation blocks; for instance, in a stream containing an aqueous leach solution and solids, two substreams are likely to be defined, one for the solution and one for the solids. To describe the different types of material flow, there are currently three types of sub stream in ASPEN PLUS: a MIXED type to represent the flow of conventional components in both chemical and phase equilibrium: CISOLID to represent the flow of solid conventional components which are considered to be inert with respect to phase equilibrium but not chemical equilibrium calculations:

a non-conventional substream type, NC, which is used to represent the flow of non-conventional components. Fig.A5 a-c show the stream data for each type of substream. Important features to note are that there are component flow locations in each substream in a simulation for each component of the substream type; thus each MIXED and CISOLID substream will carry a molar flowrate for every conventional component in the simulation and each NC substream will have a mass flowrate for each non-conventional component. All information in the execution of an ASPEN PLUS simulation is carried in SI units; although the user may use almost any unit in the input file, the input translator will convert this into SI. Data such as substream temperature, pressure and density refer to the overall substream value rather than any part of it; thus in a solid substream it is assumed that all particles are homogeneous with respect to these substream properties. If this is inappropriate then it would be necessary to have more than one NC substream, a situation which ASPEN PLUS can handle; as is described in chapter 5 it is also possible to calculate properties, such as density, from component attribute and substream attribute information for heterogeneous particles.

Distributions associated with the bulk material represented by a substream, rather than with individual components, are described by means of substream attributes. A substream attribute is a set of data that represents some characteristic of a substream in addition to the component flows, temperature etc, included in the basic substream data. In the commercial version of ASPEN PLUS there is only one type of substream attribute, particle size distribution, although, as has been described in chapter 5, two more, particle size versus density and particle size mineral fraction, were added to extend the solids processes that can be simulated; again due to the modular and flexible nature of the plex data structure, it is possible to add more as the need arises, it should be remembered that for each substream attribute, or indeed substream type,

SUBSTREAM TYPE MIXED

Conventional Component
Molar Flows (kmol/s)
Total Molar Flow (kmol/s)
Temperature (K)
Pressure (Pa)
Enthalpy (J/kg)
Molar Vapour Fraction
Molar Liquid Fraction
Entropy (J/kg/K)
Density (kg/m^3)
Molecular Wt (kg/kmol)

Fig. A5a

SUBSTREAM TYPE NC

Nonconventional Component
Mass Flows (kg/s)
Total Mass Flow (kg/s)
Temperature (K)
Pressure (Pa)
Enthalpy (J/kg)
Vapour Fraction
Liquid Fraction
Entropy (J/kg/K)
Density (kg/m^3)
Component Atrributes
Substream Attributes (if any)

Fig. A5b

SUBSTREAM TYPE CISOLID

Conventional Component
Molar Flows (kmol/s)
Total Molar Flow (kmol/s)
Temperature (K)
Pressure (Pa)
Enthalpy (J/kg)
Entropy (J/kg/K)
Density (kg/m^3)
Molecular Wt (kg/kmol)
Component Attributes (if any)
Substream Attributes (if any)

Fig. A5c

it is necessary to, at the minimum:-

a) change all unit operation blocks likely to handle the new structure

b) write stream handling subsystem and unit operation utility subroutines to ease the handling of the new substream structure

c) insert new tables into the System Definition File to allow the Input Translator to recognize the pertinent input data

d) write a report writing subroutine for the structure

Each substream attribute has two kinds of data associated with it. The first, called descriptive data, is data required to fully define an attribute type; this is left free for the user to specify so that the attribute can be tailored to the requirements of a particular simulation, although default values can be assumed. Descriptive data does not vary from stream to stream and is not carried as stream data, but rather, is stored in its own area of the simulation data structure; in the particle size distribution substream attribute, the descriptive data includes the number of intervals and their size limits. The second kind of data varies from stream to stream and is therefore carried as stream data; in the particle size distribution example this is the fraction of the substream in each size interval.

The constituent "building blocks" of an ASPEN PLUS stream are the components, the substreams and their respective attributes; these "building blocks" may be combined in many different ways to suit the requirements of a particular simulation. As a result the ASPEN PLUS stream structure is not fixed but, within a general stream structure, may take a wide variety of forms depending on the number and types of substreams, substream attributes, components and component attributes. A number of commonly occuring stream and substream structures have been

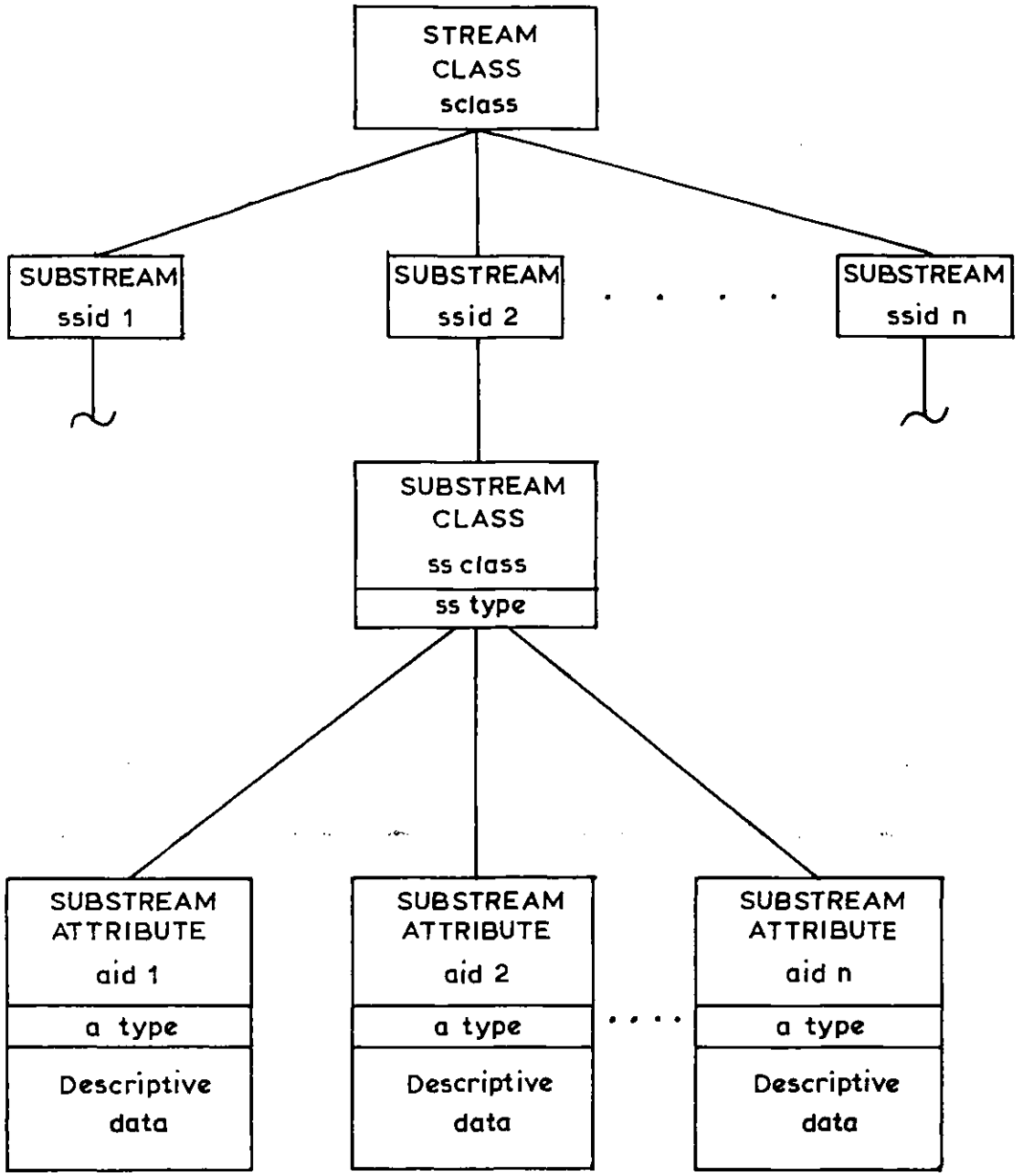


Figure A6 General stream class structure and hierarchy

built in to the system as stream or substream classes although the user may create his own stream classes through input statements.

The general structure of a stream class is shown in Fig.A6; as can be seen it is defined by a hierarchy of structures at four levels with corresponding terminology at each level; a user can define new structures or over-ride built in structures at any of these four levels. The corresponding levels and terminology are:-

STREAM CLASS: a data structure for streams consisting of one or more substreams and identified by its name "sclass"; each stream in a simulation belongs to a particular stream class, though not necessarily the same one.

SUBSTREAM: data used to represent a portion of a stream that is to be treated in a particular way by unit operation models; it is fully defined when its class has been specified and it has been given an ID label. There can be several substreams of a given class, each with its own ID.

SUBSTREAM CLASS: a data structure consisting of an area for the basic substream data (flowrates, temperature etc) and, optionally, one or more substream attributes. Each substream must belong to a defined substream class, each class being identified by a name "ssclass"; each class must belong to a certain substream type, "sstype". Substream types are part of the ASPEN PLUS system defined by the SDF, but substream classes may be defined by the user.

SUBSTREAM ATTRIBUTE: data that represents a characteristic of the substream in addition to the component flows, temperature etc represented by the basic substream data; they are typically used to represent physical characteristics of a substream such as a particle size distribution and/or density distribution. It is important to note that substream attributes are associated with, and as such are a property of, the bulk material represented by the substream, not the individual components of the substream

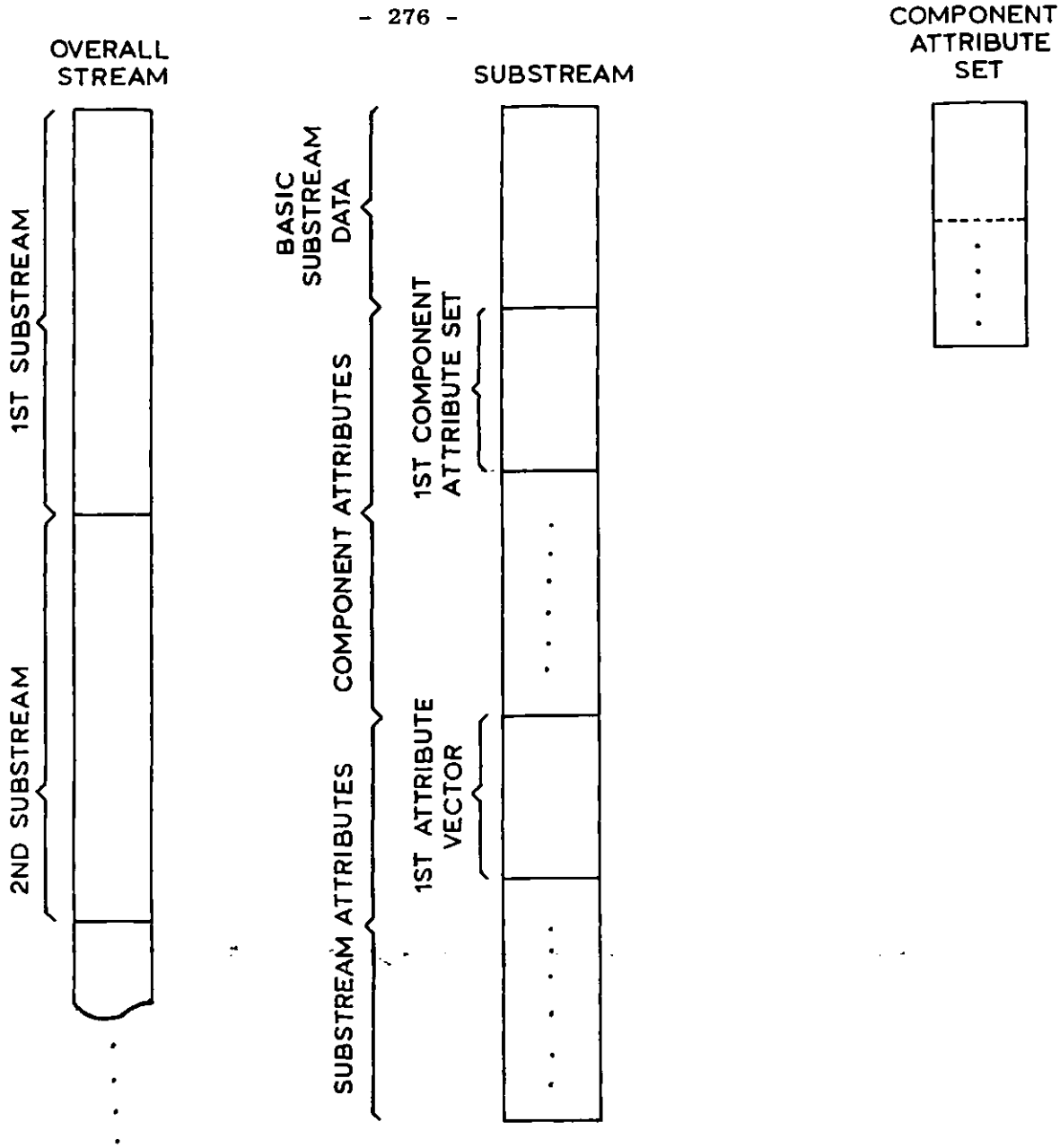


Figure A7 General Stream Vector Structure

(NB. substream attribute descriptive data not shown)

	5.55	Water flow
	5.55	Total flow
	300.0	Temperature
	1.0	Pressure
MIXED SSTR.	0.0	Enthalpy
	0.0	Molar vapour fraction
	0.0	Entropy
	1000.0	Substream density
	18.0	substream mol. wt.
	1500.0	Ore flow
	1500.0	Total flow
	300.0	Temperature
	1.0	Pressure
NCPSD SSTR.	0.0	Enthalpy
	0.0	molar vapour fraction
	0.0	Entropy
	1.0	Substream density
	1.0	Substream mol. wt.
	3000.0	
COMPONENT ATTR.	⋮	
	0.15	
	0.15	
	0.10	
	0.20	
SSTR. ATTRIBUTE	0.1	
	0.1	
	0.06	
	0.08	
	0.06	

Figure A8 Stream S1 (Fig. A4) data vector (descriptive data not shown)

not the individual components of the substream. There can be several attributes of a given type, each with its own ID and descriptive data.

The general form of the stream vectors generated by the ASPEN PLUS system is shown in Fig.A7, whilst Fig. A8 illustrates this using as an example, the stream S1 generated by the stream statements in Fig.A4.

Unit Operation Blocks

An ASPEN PLUS block refers to the process flowsheet element representing a process unit; this block comprises of three basic subroutine groupings centred around a model subroutine. The model subroutine is used to describe the unit operation and define the transformation that takes place in converting input streams into output streams. The primary function of a unit operation block during conveyance of the heat and material balances is to calculate output stream variables given the input stream variables; on each such executiion of a block, information may be written onto a history file to provide a record of the progress made towards convergence of the simulation, the amount of information written being controlled by a user defined diagnostic level, following the convergence of the heat and material balance, the block is accessed for a subsequent "results pass" to calculate results which are not needed for convergence of the flowsheet heat and material balance, but which are of interest in the final report of the simulation. After simulation of the process is complete the block is accessed for a final time during the report writing stage; a major part of the simulation report is written by unit operation block subroutines, the information being written to the file consisting of five standard sections plus optional special sections. The five standard sections are the header, which includes block ID, model name, inlet and outlet stream ID's, block description and block status: the component material balance: total material and energy balance: input specifications and default values: results, including all useful information not readily available as stream results. If an ASPEN PLUS simulation is to include equipment

costing and economic evaluation as well as material and energy balancing, the block will be accessed once more, between the results and report passes; in this stage it is necessary to, where necessary, carry out equipment sizing calculations and to write equipment size information into the appropriate array locations for retrieval by the costing subroutines.

The way in which unit operation blocks carry out their various roles is perhaps best illustrated by Fig. A9, which shows the flow of information to and from a block. As is indicated by Fig. A9, a unit operation block consists of an SDF table and three subroutine groupings, the interface, model and report routines, each of which are documented as templates in the on-line system documentation.

The SDF block table, in common with other table types, defines the block bead structure and input language; it also defines input and output stream restrictions, special report options and model design specification variable types. The basic structure of the block bead is shown in Fig.A11 which illustrates its variable length, this flexibility being made possible and defined by the input translator through the use of the model SDF table; not shown in Fig. A10 are the fortran commons WORK and IWORK which contain storage space for use by the unit operation block and accessed through location pointers in the IWDIR array. There are two major types of variable array location used by an ASPEN PLUS block other than the fixed arrays such as run control parameters and the report option vector. The first type of array has storage locations which are maintained throughout a simulation and which are used for a specific array element or variable; these are stored in the REAL and INT locations in the block bead. These locations are used for input parameters, block results and variables which are needed in subsequent accessing of the block; this last classification includes variables such as stream parameters which remain constant during the simulation and restart parameters which are used to accelerate model

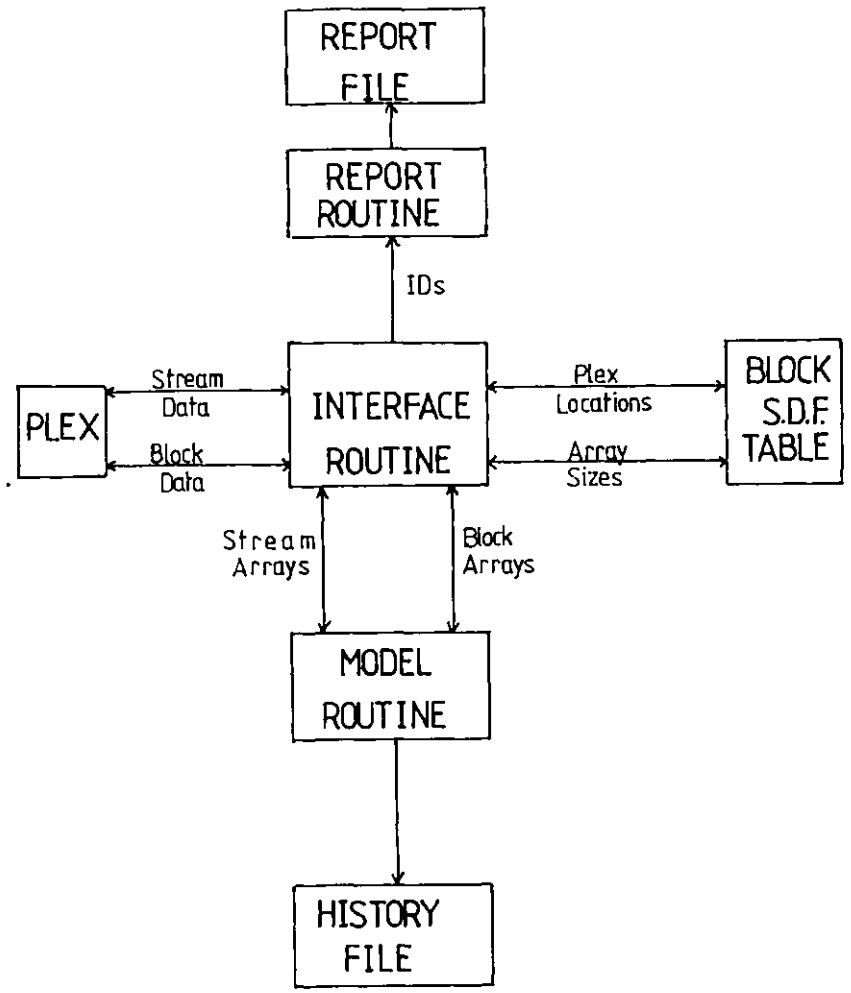


Figure A9 Block Information flow

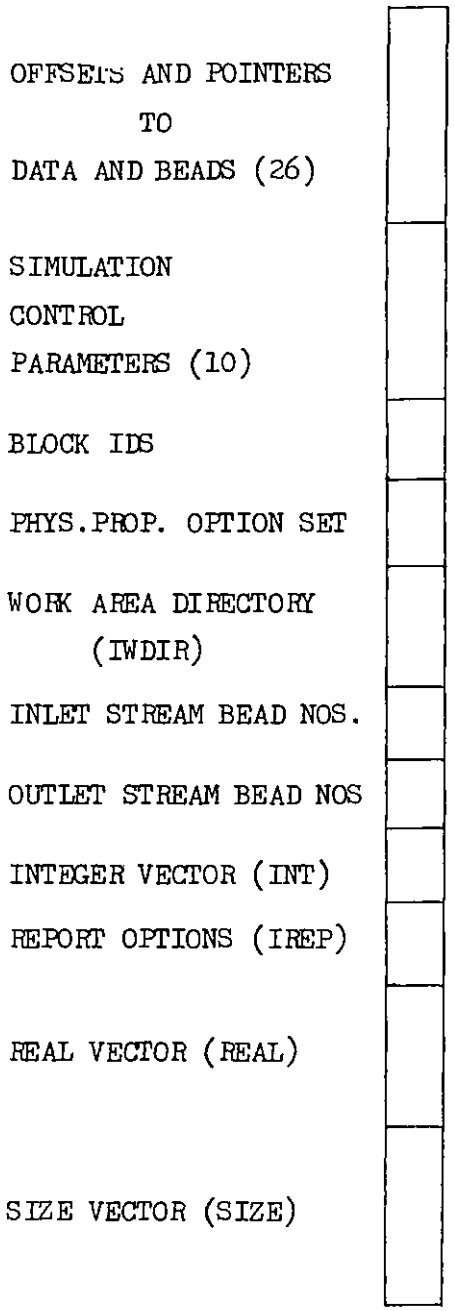


Figure A10 BLOCK Bead

convergence the next time that the model is accessed. The arrays are defined in the SDF by secondary and tertiary keywords, these being the same keywords as used in the input file, and a code system which is processed during the input translation stage to determine the bead dimensions. The location and dimensions of both INT and REAL arrays are stored in the initial elements of the INT section of the block bead. The second type of array location is used for block variables which are only called during each block execution; examples of these are stream flowrates and block internal variables. These arrays use a common location, WORK, which is used by all blocks in the process; thus any WORK location from which part of the block takes data values, should have had those values written into it during that block execution. Starting locations for each array location are stored in the IWDIR array of the block bead. The amount of WORK common space is determined by entries in a work area table in the SDF block table, these entries using similar codes to the keyword entries to determine dimensions.

The block interface routine is the subroutine which is called by the main simulation program; its initial purpose is to determine the stage which program execution has reached and to start the relevant internal calculation procedures. Other than acting as a calling subroutine, the interface routine's main role is in decomposing the block and stream beads and block work area into their relevant sub-arrays and to transform the bead data structure of these subarrays into data structures which can be more efficiently managed during the model calculations. In carrying out this role the interface routine generally calls on subroutines and does not actually pass actual data but rather only array addresses. When the model subroutine grouping is called it should not need to access the plex directly, except perhaps for individual data entries, and the arrays which the model acts on should already have been set up. Once model execution is complete it is also the interface routine which writes the results back into the relevant plex locations. The model interface routines make extensive

use of two suites of subroutines and functions, the Stream Handling Subsystem and Unit Operation Utilities.

Having been presented with all of the relevant data, it is the role of the model group of subroutines to perform the transformations on that data to predict the performance of the unit operation and determine the outlet stream values. There are no ASPEN PLUS conventions for the structure of the model subroutines, other than the options of ignoring certain calculations until the results pass and of writing information on the progress of the simulation calculations into the History File. Excepting these and the model subroutine argument list, it would be possible to transplant a stand-alone unit operation model program into this subroutine grouping.

Whilst the duties that the block report writer is expected to perform may seem complex, due to the drafting of a template and the provision of utility subroutines, the development of a report writer for a unit operation block is not arduous. The additions to the template to be made include printing stream IDs, writing statements for errors made, and printing input and results information. The amount of results information depends on the model developer and what information is needed in addition to stream data; if the possible required information is considerable it is possible to insert optional special report sections.

Flowsheet Convergence and Control

ASPEN PLUS uses an advanced sequential modular (427) approach to flowsheet convergence. In this approach, individual unit operation models compute outlet streams and block results, given inlet streams and block parameters. Recycle streams are born and the tear stream variables are used as iteration variables. Design specifications are imposed by freeing certain block parameters or feed stream variables and adjusting their value to meet specifications on stream variables or block results.

The first step in structuring the computations for sequential modular solution, whether done manually or automatically, is to partition the

flowsheet into "maximal cyclic subsystems": these are sections of the flowsheet that are to be converged prior to any computation of down stream subsystems. Next a set of tear streams is selected to break the recycle loop; the choice is aimed at achieving the best convergence properties of the resulting algorithm. Loops are then nested by dividing iteration variables into subsets of variables to be converged simultaneously and specifying the order in which they are to be converged. Finally the computational sequence is determined.

In many process simulators the user is responsible for all structuring of computations and specifies the computational sequence directly. In ASPEN PLUS the system is capable of complete automatic determination of the computational sequence, but the user can impose as many constraints as desired, from selecting certain tear streams to specifying the entire sequence.

In ASPEN PLUS there are no differences between the convergence of recycle streams and the satisfaction of design specifications; both are handled using convergence blocks. In recycle stream convergence the stream variables are adjusted to drive the difference between assumed and recycle variables to zero, whilst in satisfying design specifications the intention is to drive the difference between the required and calculated values of design specification to zero. There are a number of convergence block types available ranging in complexity from direct substitution to Broyden's method. Using the input language, the user can access any flowsheet variable and carry out any transformation of these variables by entering program statements or subroutines in FORTRAN. Accessed variables, or transformations of them may be used as design specifications to be driven to zero by the convergence blocks.

Cost Estimation and Economic Evaluation

The purpose of the ASPEN PLUS Cost Estimation and Economic Evaluation Subsystem is to calculate the profitability of a simulated process. From

details of equipment sizes and process data the subsystem can estimate the total required capital investment, annual operating expenses and revenues, and measures of the process profitability. These details can be either user-supplied, with the subsystem being used in a stand-alone mode, or can be obtained directly from the results of an ASPEN PLUS simulation run.

The defined accuracy of the capital cost investment estimation is 30%, although, due to the possible incorporated detail of the factorial methods used, it is likely that more reliable estimates are obtainable. A program module, of similar structure to the simulation blocks, is provided for each equipment class to calculate the purchased cost of each piece of equipment. The purchased cost is factored, on a unit operation, sectional or plant basis, to provide a battery-limit estimate of the material costs and labour hours involved. Further factors are then applied to these figures to determine the required capital investment.

Operating costs are estimated, again factorially, based on the cost of consummables, operating labour hours and the capital investment. These are combined with the estimates of capital requirements and revenues to calculate annual cash flows, which are then used to provide details of the economic value of the project, using both discounted and non-discounted cash flow techniques. All of the calculation procedures are part of the subsystem although it is a fairly easy matter for new costing modules to be added to the system to extend its application. The only potential drawbacks to the procedures from a minerals extraction viewpoint are the inability to calculate the cost of reagent additions, though this may be rectified by breaking them as utilities in an analagous fashion to power and water, and the inability to allow for price fluctuations according to product composition.

Appendix 2: MECON/PLANT Cost Estimation and Economic Evaluation System Documentation

The basic structure and calculation procedures of the MECON/PLANT system have been described in Chapter 5.2 of this thesis; in this Appendix the input file and the unit operation cost blocks are described.

As was explained in Chapter 5.2, the data file PROFLOW is used to input the basic cost data to the MECON program and calculate the capital and operating costs, and revenues; the only changes to be made to this data file should occur when the project life and/or starting data is to be altered. It is assumed that plant construction occurs during the first calendar year and the operating life after this time is defined in data set 10. The main input to the system is through another data file which is read by the PLANT program, which uses the information to calculate the sectional capital and operating costs and feeds the necessary information into the PROFLOW data file. The complete input file is shown in Table A2.1; a rigid format of eight characters, including spaces, followed by a real number (format F20.8) is followed except in indicating the start and end of a cost section when only one word is used. Whilst not particularly flexible and potentially resulting in a long list of variable names and values, this format was chosen as being fairly robust, easy to follow and allowing extensive use of default values, especially for cost factors. In Table A2.1 those variable names which are underlined must be included in the order shown, whilst the other variables may take any order, if they appear at all, between the preceding and succeeding compulsory variable. Each section may only be called up to once in a program run, though the unit operation cost blocks may be called more than once, with no maximum order set.

The unit operation cost blocks each calculate the purchased cost of the equipment and, where necessary, the power and major consummables requirements per shift. The power requirements are estimated using the total installed

motor ratings whilst the consumption of media and reagents is calculated based on the input of unit consumption rates and volume flowrates or installed power. Most equipment purchase costs are calculated using cost relationships published by the CIMM (409); the exceptions are the copper electrowinning block, which is based on Balberyszski and Anderson's work (412). All cost data is in Canadian dollars and has been converted to a common base Marshall and Swift Cost Index (Mining and Milling) value of 800. The equipment listings assumed in each block are as follows:-

Gyratory Crusher

Crusher (incl drive and lubrication system)

Motor

Cone Crusher

Crusher (incl drive and lubrication system)

Motor

Vibrating Screen

Screens (incl.drive and feed bse)

Motor (20 h.p.)

Screen Cloth

Conveyor Belt

Belt (incl. structure and all ancillary equipment)

Ore Bin

Bin

Ball Mill

Mill (incl. liners, gearing and lubrication system)

Motor

Ball Charge (50% of internal mill volume)

Hydrocyclone Set

Cyclones (incl.fittings and liners)

Pump and Motor

Flotation Bank

Cells (incl. motors, feed boxes, optional paddles and blowers)

Conditioner (optional)

Concentrate pump and motor

Tailings " " "

Filter

Filter

Thickener and CCD

Thickener (incl. rakes, drives, motors and walls)

U/F Pump and Motor

Leach Cascade

Tanks

Agitator mechanisms and motors

Solvent Extraction Bank

Mixer bank, agitator and motor

Settler assembly

(NB. excludes solvent inventory which may be inputted using section factors or costs).

Electrowinning

Complete tankhouse and inventory

Tailings Pond

Pond, dam and pumps

INPUT NAME	DEFAULT VALUE	VARIABLE EXPLANATION
M+S INDEX		M+SA (MINING + MILLING COST INDEX)(BASE=800)
CLABRATE	(17.5)	AVERAGE HOOURLY PAYRATE OF CONSTRUCTION LABOUR
CLABPROD	(1.0)	PRODUCTIVITY OF CONSTRUCTION LABOUR
MFSERBLD	(0.075)	SERVICE BUILDING MATERIAL FACTOR
LFSERBLD	(0.03)	" " LABOUR "
MFSERSYS	(0.06)	SERVICE SYSTEMS MATERIAL FACTOR
LFSERSYS	(0.02)	" " LABOUR "
MFDIRADO	(0.0)	DIRECT ADDITIONAL MATERIAL FACTOR
LFDIRADD	(0.0)	" " LABOUR "
MFSITDEV	(0.01)	SITE DEVELOPMENT MATERIAL FACTOR
LFSITDEV	(0.002)	" " LABOUR "
FFREIGHT	(0.02)	FREIGHT COST FACTOR
MFINDFLD	(0.27)	INDIRECT FIELD MATERIAL FACTOR
LFINDFLD	(0.73)	" " LABOUR "
FPROJMAN	(0.06)	PROJECT MANAGEMENT FACTOR
FENGHOFF	(0.12)	ENGINEERING + HOME OFFICE FACTOR
FFEESETC	(0.08)	FEES, PERMIT + INSURANCE FACTOR
FPROCONT	(0.2)	PROCESS CONTINGENCY FACTOR
FPRODEFF	(0.1)	PROJECT DEFINITION FACTOR
FLAND	(0.02)	LAND FACTOR
CLAND	(0.00)	LAND COST
FROVALTY	(0.00)	ROYALTY FACTOR
CROVALTY	(0.00)	ROYALTY COST
FEEDTONS		ORE MILLED (TONNES/WEEK)
FEED COST		ORE COST ($/WEEK)
NO.PROD		NUMBER OF PRODUCTS (MAX.5)
PRICE)	for each	PRODUCT REVENUE ($/KG)
AMOUNT)	product	PRODUCT AMOUNT (KG/WEEK)
CCSTEEL	(0.0)	PRICE CRUSHING LINERS ($/KG)
CGSTEEL	(0.0)	" GRINDING BALLS ($/KG)
CFLOTREA	(0.0)	" FLOTATION REAGENTS ($/KG)
CACID	(0.0)	" ACID ($/KG)
CSOLVENT	(0.0)	" SX REAGENT ($/KG)
CFLOC	(0.0)	" FLOCCULANT ($/KG)
CPOWER		" POWER ($/KWH)
CWATER		" WATER ($/$M^3$)
FSUPRVIS	(0.2)	SUPERVISION COST FACTOR
FADMIN	(0.6)	ADMINISTRATION COST FACTOR
FFRING	(0.4)	FRINGE BENEFIT " "
FFMANLAB	(0.036)	MAINTENANCE LABOUR COST FACTOR
FSMAINUP	(0.024)	" SUPPLY " "
FSPSUP	(0.06)	OPERATING " " "
FSTATTAX	(0.02)	LOCAL TAX COST FACTOR
FSTATINS	(0.0075)	LOCAL INSURANCE COST FACTOR
FOTHER	(0.03)	EXTRA OPERATING " "
OPLABPAY		OPERATING LABOUR BASIC PAY ($/HR)
NO.WEEKS		NUMBER OF WEEKS OPERATED ANNUALLY

TABLE A2.1: MECON/PLANT Input File

The remainder of the input consists of the section and unit operation information; each section may only be used once or not at all, but each unit operation block may be called any number of times. For purposes of illustration the input formats for all sections and unit operations are listed; although not underlined, every line of a unit operation block input must be included in the data file. The end of each section is indicated by the keyword SECTION and the end of the data file by the key wordEND. The format for inputting cost factors, always located between SHIFTS AND OPERATOR, is identical for each section; a table of the various default values is given in table A2.2.

```
CRUSHING                                    CRUSHING SECTION INPUT
                                                OPERATED
SHIFTS                                      NO. OF SHIFTS PER WEEK
MFOVERAL      (0.0)          OVERALL INSTALLATION MATERIAL FACTOR
LFOVERAL      (0.0)             "          "       LABOUR      "
CMOVERAL      (0.0)             "          "       MATERIAL COST
HLOVERAL      (0.0)             "          "       LABOUR HOURS
MFPIPING      (   )          PIPING MATERIAL FACTOR
LFPIPING      (   )             "   LABOUR      "
CMPIPING      (0.0)             "   MATERIAL COST
HLPIPING      (0.0)             "   LABOUR HOURS
MFFOUND       (   )          CONCRETE MATERIAL FACTOR
LFFOUND       (   )             "   MATERIAL COST
CMFOUND       (0.0)             "   MATERIAL COST
HLFOUND       (0.0)             "   LABOUR HOURS
MFSTEEL       (   )          STRUCTURAL MATERIAL FACTOR
LFSTEEL       (   )             "   LABOUR         "
CMSTEEL       (0.0)             "   MATERIAL COST
HLSTEEL       (0.0)             "   LABOUR HOURS
MFINSTR       (   )          INSTRUMENT MATERIAL FACTOR
LFINSTR       (   )             "   LABOUR FACTOR
CMINSTR       (0.0)             "   MATERIAL COST
HLINSTR       (0.0)             "   LABOUR HOURS
MFELEC        (   )          ELECTRICAL MATERIAL FACTOR
LFELEC        (   )             "   LABOUR FACTOR
CMELEC        (0.0)             "   MATERIAL COST
HLELEC        (0.0)             "   LABOUR HOURS
MFINSUL       (   )          INSULATION MATERIAL FACTOR
LFINSUL       (   )             "   LABOUR         "
CMINSUL       (0.0)             "   MATERIAL   COST
HLINSUL       (   )             "   LABOUR HOURS
MFPAINT       (   )          PAINTING MATERIAL FACTOR
LFPAINT       (   )             "     LABOUR     "
CMPAINT       (   )             "     MATERIAL COST
HLPAINT       (0.0)             "     LABOUR HOURS
MFMISC        (   )          MISCELLANEOUS MATERIAL FACTOR
LFMISC        (   )             "          LABOUR      "
CMMISC        (0.0)             "          MATERIAL COST
HLMISC        (0.0)             "          LABOUR HOURS
```

LFINSTAL	( )	INSTALLATION LABOUR FACTOR
HLINSTAL	(0.0)	" " HOURS
MFBUILD	( )	BUILDING MATERIAL FACTOR
LFBUILD	( )	" LABOUR "
CMBUILD	(0.0)	" MATERIAL COST
HLBUILD	(0.0)	" LABOUR HOURS
MFSPARE	(0.0)	CAPITALIZED SPARES FACTOR
CMSPARE	(0.0)	" " COST
MFTEST	(0.0)	TESTING MATERIAL FACTOR
LFTEST	(0.0)	" LABOUR "
CMTEST	(0.0)	" MATERIAL COST
HLTEST	(0.0)	" LABOUR HOURS
MFADD	(0.0)	ADDITIONAL MATERIAL FACTOR
LFADD	(0.0)	" LABOUR "
CMADD	(0.0)	" MATERIAL COST
HLADD	(0.0)	" LABOUR HOURS
MFEXTRA	(0.15)	UNLISTED ITEMS MATERIAL FACTOR
CMEXTRA	(0.0)	" " " COST
OPERATOR		NUMBER OF OPERATORS ON SECTION
GYRATORY		GYRATORY CRUSHER COST BLOCK
NUMBER		NUMBER OF CRUSHERS IN PARALLEL
SIZE 1		RECEIVER OPENING (M)
SIZE 2		MANTLE DIAMETER (M)
MOTOR		MOTOR SIZE (KW)
ABRASION		ABRASION RATE (KG/KWH)
CONE		CONE CRUSHERS COST BLOCK
NUMBER		NUMBER OF CRUSHERS IN PARALLEL
DIAMETER		DIAMETER OF DISCHARGE ANNULUS (M)
MOTOR		MOTOR SIZE (KW)
ABRASION		LINER ABRASION RATE (KG/KWH)
SCREEN		VIBRATING SCREEN COST BLOCK
NUMBER		NUMBER OF SCREENS IN PARALLEL
WIDTH		SCREEN WIDTH (M)
LENGTH		" LENGTH (M)
DECKS		NUMBER OF DECKS (1.0,2.0 or 3.0)
CLOTH		DECKING MATERIAL (1.0,2.0 or 3.0)
BINS		ORE BIN COST BLOCK
NUMBER		NUMBER OF BINS
STORAGE		WEIGHT CAPACITY OF EACH BIN (KG)
BULKDEN		MATERIAL BULK DENSITY (KG/M^3)
CONVEYOR		BELT CONVEYOR COST BLOCK
NUMBER		NUMBER OF BELTS IN PARALLEL
WIDTH		BELT WIDTH (M)
LENGTH		BELT LENGTH (M)
SECTION		
GRINDING		GRINDING SECTION INPUT
SHIFTS		NUMBER OF SHIFTS OPERATED WEEKLY
OPERATOR		NUMBER OF OPERATORS ON SECTION
BALLMILL		BALL MILL COST BLOCK
NUMBER		NUMBER OF MILLS IN PARALLEL
DIAMETER		MILL DIAMETER (M)
L/D		LENGTH TO DIAMETER RATIO
MOTOR		MILL MOTOR SIZE (KW)
ABRASION		BALL + LINER ABRASION RATE
HCYCLONE		HYDROCYCLONE COST BLOCK
SETNO		NUMBER OF CYCLONE SETS
CYCNO		NUMBER OF CYCLONES IN EACH SET
DIAMETER		CYCLONE DIAMETER (M)

Table A2.1 (cont)

VOLFLOW	PULP VOLUME FLOWRATE TO EACH SET (M^3/SEC)
PUMPMOTR	PUMP MOTOR SIZE
SECTION	
FLOAT	FLOTATION SECTION INPUT
SHIFTS	NUMBER OF SHIFTS OPERATED WEEKLY
OPERATOR	NUMBER OF OPERATORS ON SECTION
FLOTBANK	FLOTATION BANK COST BLOCK
BANKNO	NUMBER OF BANKS IN PARALLEL
CELLNO	NUMBER OF CELLS IN A BANK
CELLVOL	VOLUME OF CELLS
CELLMOT	CELLMOTOR SIZE (KW)
CONRES	CONDITIONER RESIDENCE TIME (SEC)
BLOWER	BLOWER OPTION (0.0 or 1.0)
PADDLE	PADDLE OPTION (0.0 or 1.0)
TAILVOL	TAILING VOLUME FLOWRATE (M^3/SEC)
CONCVOL	CONCENTRATE " " "
TAILMOT	TAILING PUMP MOTOR SIZE (KW)
CONCMOT	CONCENTRATE PUMP " " "
REAGENT	REAGENT CONSUMPTION (KG/M/SEC
SECTION	
SLSEP	SOLID/LIQUID SEPARATION SECTION INPUT
SHIFTS	NUMBER OF SHIFTS OPERATED WEEKLY
OPERATOR	NUMBER OF OPERATORS ON SECTION
FILTER	FILTRATION COST BLOCK
HORZBELT	HORIZONTAL BELT) name of filter
ROTYDISC	ROTARY DISC) type. (only 1 in
ROTYDRUM	ROTARY DRUM) each block).
NUMBER	NUMBER OF FILTERS
AREA	FILTER AREA (M^2)
COAT	OPTIONAL COATING FOR WETTED PARTS
VACPRES	VACUUM PRESSURE (IN.HG)
VACCAP	VACUUM PUMP CAPACITY (M^3/SEC)
VACMOT	VACUUM PUMP MOTOR SIZE (KW)
VACNO	NUMBER OF VACUUM PUMPS
THICKNER	THICKENER COST BLOCK
AREA	SETTLING AREA
FLOCUSE	FLOCULLANT CONSUMPTION (KG/M^3)
FEEDVOL	FEED PULP VOLUME FLOWRATE (M^3/SEC)
U/F VOL	UNDERFLOW PULP VOLUME FLOWRATE (M_3/SEC)
PUMPMOTR	" PUMP MOTOR SIZE (KW)
TAILINGS	TAILINGS POND COST BLOCK
CAPACITY	POND CAPACITY (KG/SEC)
CONCHAND	CONCENTRATE STORAGE COST BLOCK
CAPACITY	CONCENTRATE PRODUCTION (KG/SEC)
SECTION	
LEACH	LEACH + CCD SECTION INPUT
SHIFTS	NUMBER OF SHIFTS OPERATED WEEKLY
OPERATOR	NUMBER OF OPERATORS ON SECTION

Table A2.1 (cont).

LEACH	LEACH COST BLOCK
NUMBER	NUMBER OF LEACH TANKS
FEEDFLOW	FEED FLOWRATE (M^3/SEC)
RES.TIME	RESIDENCE TIME (SEC)
ACID.CON	ACID CONSUMPTION (KG/SEC)
CCD	COUNTER-CURRENT DECANTATION COST BLOCK
NUMBER	NUMBER OF STAGES
AREA	SETTLING AREA OF EACH STAGE
FLOCUSE	FLOCULLANT CONSUMPTION (KG/M^3)
FEEDVOL	FEED PULP VOLUME FLOWRATE (M^3/SEC)
U/F VOL	UNDERFLOW " " "
U/F MOT	" PUMP MOTOR SIZE (KW)
SECTION	
SX-EW	SOLVENT EXTRACTION - ELECTROWINNING SECTION
SHIFT	NUMBER OF SHIFTS OPERATED WEEKLY
OPERATOR	NUMBER OF OPERATORS
SX	SOLVENT EXTRACTION COST BLOCK
NUMBER	NUMBER OF STAGES IN BANK
MIXERVOL	VOLUME OF MIXER (M^3)
SETTLVOL	VOLUME OF SETTLER
MIXMOTOR	SIZE OF MIXER MOTOR (KW)
SOLVLOSS	LOSS OF SOLVENT (KG/SEC)
EW	ELECTRO-WINNING COST BLOCK
CAPACITY	COPPER PRODUCTION RATE (KG/YEAR)
CELL VOLT	CELL VOLTAGE DROP (V)
CURRDENS	CURRENT DENSITY (A/M^2)
ELECAREA	ELECTRODE AREA (M^2)
ACIDCONS	ACZD CONSUMPTION (MG/SHIFT)
SECTION	
END	

SECTION

FACTOR	CRUSHING	GRINDING	FLOTATION	SLSEP	LEACH-CCD	SX-EW
MFPIPING	0.06	0.14	0.25	0.25	0.335	0.335
LFPIPING	0.0017	0.0036	0.0048	0.0048	0.0080	0.0093
MFFOUND	0.04	0.045	0.0515	0.0515	0.0525	0.0525
LFFOUND	0.0031	0.003	0.0035	0.0035	0.0035	0.004
MFSTEEL	0.03	0.03	0.040	0.040	0.045	0.045
LFSTEEL	0.0009	0.001	0.001	0.001	0.0012	0.0013
MFINSTR	0.02	0.03	0.03	0.03	0.045	0.045
LFINSTR	0.0005	0.0007	0.006	0.0008	0.0008	0.001
MFELEC	0.05	0.052	0.075	0.075	0.085	0.085
LFELEC	0.0021	0.0024	0.0028	0.0028	0.0033	0.0036
MFINSUL	0.02	0.02	0.015	0.015	0.01	0.01
LFINSUL	0.0017	0.0015	0.012	0.0012	0.0009	0.0009
MFPAINT	0.003	0.003	0.0035	0.0035	0.0035	0.0035
LFPAINT	0.0006	0.0006	0.0006	0.0006	0.0006	0.0006
MFMISC	0.03	0.03	0.035	0.035	0.035	0.035
LFMISC	0.0014	0.0014	0.0018	0.0018	0.0021	0.0021
LFINSTAL	0.017	0.0188	0.0187	0.0187	0.0167	0.0172
(MFOVERAL	0.253	0.290	0.500	0.500	0.611	0.611)
(LFOVERAL	0.029	0.033	0.035	0.035	0.0371	0.040)
MFBUILD	0.05	0.055	0.055	0.055	0.050	0.060
LFBUILD	0.003	0.003	0.003	0.003	0.003	0.003

Table A2.2: Default installation factor values for the 6 process sections.

(MFOVERAL + LFOVERAL are not default values but are included to indicate the total factors).

References

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