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RENISON LIMITED

THE METALLURGICAL TREATMENT OF BLUE TIER MINERALISATION

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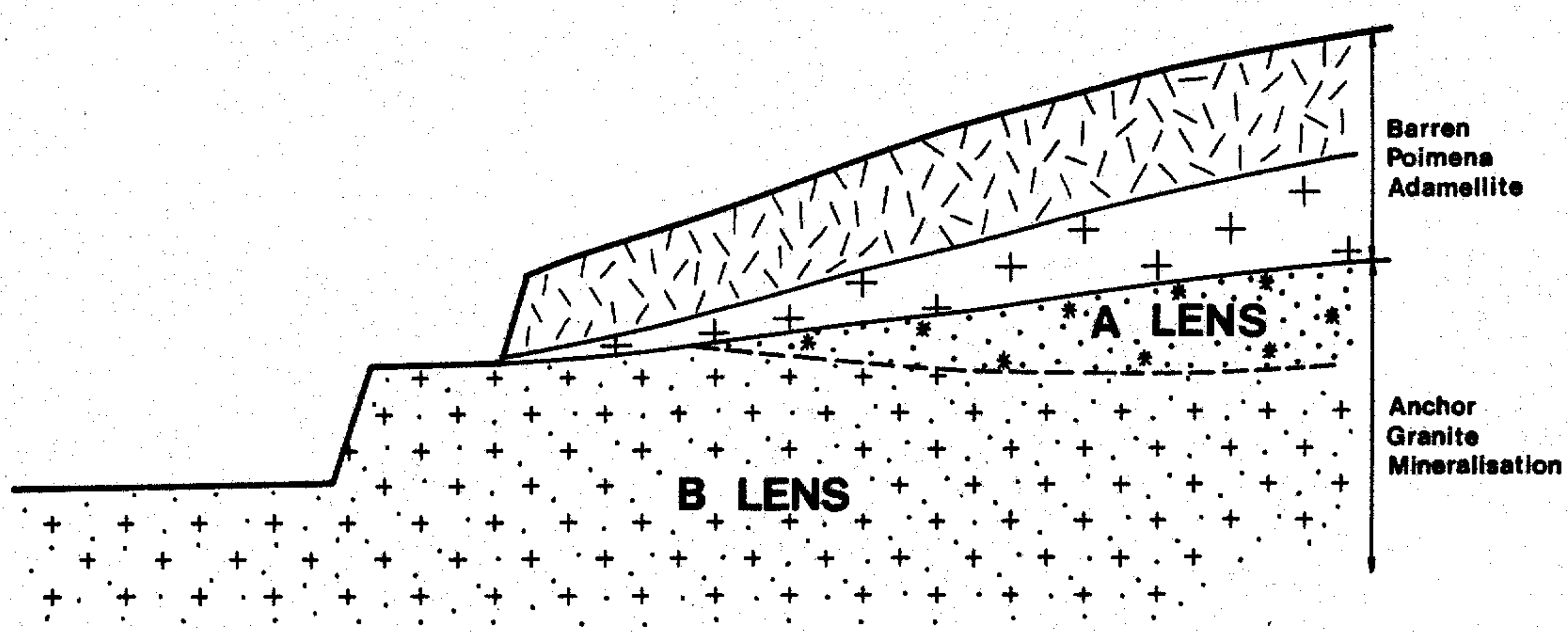
1. OBJECTIVE

To define a suitable flowsheet for the processing of Blue Tier mineralisation.

FIGURE 1



SKETCH – SECTION THROUGH ANCHOR TIN DEPOSIT

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2. SUMMARY

In October 1979, following a program of diamond drilling on the Anchor Mine, a geological reserve of 2,000,000 tonnes of 0.40 per cent tin (as cassiterite) was calculated using a 0.2 per cent tin cut-off grade. On the basis of this information, an Indicative Feasibility Study was prepared to assess the potential viability of a mining operation in light of the current data. Following the I.F.S. it was decided that further tonnage was required to make the deposit economically attractive and a program of further exploration was commenced.

The estimates of metallurgical performance and processing equipment requirements used in the I.F.S. were based on limited data. In order to accurately define treatment requirements and performance a metallurgical study was commenced to parallel and complement the exploration program. This report details the findings of the metallurgical study.

Exploration is continuing in the area, however, the reserve is currently estimated as 3,500,000 tonnes of 0.27 per cent tin.

A relatively simple treatment route is required to process Blue Tier mineralisation. Two stage crushing followed by single stage, closed circuit grinding would reduce the rock to a size suitable for cassiterite recovery. Cassiterite recovery would be achieved by a two stage spiral circuit followed by concentrate cleaning utilising shaking tables.

Such a circuit is anticipated to yield 89 per cent tin recovery to a concentrate containing 53 per cent tin from a feed grade of 0.27 per cent tin.

At a treatment rate of 350,000 tonnes/yr such a circuit would produce 841 tonnes of tin per year at an operating cost of $3.16 per tonne treated.

Such a concentrator would require capital expenditure of approximately 6,343,000 dollars to establish today.

A further two cases in which the annual treatment rate is 500,000 and 1,100,000 tonnes/yr were also examined. Although both cases require additional capital expenditure and result in an increase in total annual operating costs, treatment costs per tonne milled are reduced in both cases.

3. INTRODUCTION

3.1 The Deposit

The Anchor Mine is situated some 20 kilometers west of St. Helens in north-east Tasmania. The now abandoned workings occur in a series of granitic intrusive rocks known as the Blue Tier Batholith. Cassiterite mineralisation occurs in an altered zone at the top of a medium grained granite and beneath the contact of the granite and an overlying porphyritic adamellite. The deposit has been divided into two sections:

1) 'A' Lens which is generally higher grade, contains higher levels of base metal sulfides and lies to the north-east of the old workings.

2) 'B' Lens which is generally lower grade with more erratic mineralisation and lies below 'A' Lens and in the floor of the open pits.

Tin occurs as cassiterite in the mineralised zone. Sulfide traces occur within the altered zone, however, there is no unique association of the sulfides with cassiterite.

Figure 1 shows a schematic representation of a section through the deposit.

3.2 Mining the Deposit

Initial mining by open pit methods was proposed in the I.F.S. Later in the life of the mine, room and pillar and open stoping methods of underground mining would be utilised as increasing overburden depths render open pit mining uneconomic.

3.3 Size of Operation

The I.F.S. was based on a geological reserve of 2 million tonnes of 0.40 per cent tin. The continuing exploration in the area has increased the reserve to approximately 3,500,000 tonnes. However the grade of the reserve has dropped to 0.27 per cent tin. The scale of the operation, should the project become viable, is undefined although it can generally be described as small scale.

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This study has examined several cases in which the treatment rate varies. A base case of a ten year mine life resulting in a throughput of three hundred and fifty thousand tonne/yr. was chosen.

At this point it is probably pertinent to outline the underlying philosophy of this study. At all times the lowest cost (capital and operating) option has been selected, sometimes at the expense of recovery. It was felt that a low cost, reasonably efficient concentrator was more likely to be economically viable than a high cost, highly efficient concentrator. To quote from the I.F.S., "If the operation is not envisaged as a tightly run marginal mine, it will fail.".

3.4 Previous Testwork

Testwork prior to the I.F.S. was limited to heavy liquid assessments, some superpanning and jigging separations and mineralogical examinations carried out on mineralised intersections from diamond drill holes. The testwork indicated:

1) Cassiterite grainsize was relatively coarse although variable.

2) Problems may be encountered producing high grade concentrates due to the abundance of topaz and biotite.

3) Sulfides present in low concentrations were not composite with cassiterite. Sulfide assemblage was complex including sulfides of iron, copper, arsenic, zinc, bismuth and molybdenum.

4) Silver was present possibly as a sulfide and was recovered with cassiterite.

Based on this limited data a preliminary flowsheet similar to the complex Renison flowsheet was proposed and estimated to produce 85 per cent recovery of tin to a concentrate containing 55 per cent tin.

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3.5 Justification for this Study

The information upon which the I.F.S. proposals were based was limited and merely enabled an indication of the likely metallurgical performance and equipment requirements to be made. As a result the present study was commenced to enable metallurgical characteristics and performance to be confidently defined and to permit detailed processing requirements to be estimated.

3.6 Approach During this Study

The approach during this testwork has been to:

1) Further investigate and define liberation characteristics so that a grind size could be determined. Such an investigation results in an appreciation of variations in grain size from sample to sample.

2) Carry out suitable testwork to enable crushing and grinding equipment requirements to be defined. Requirements are dependent on the desired size of the product which is defined from 1) above.

3) Determine the feasibility of using heavy medium separation as a preconcentration stage.

4) Determine the most suitable means of recovering the cassiterite. The relatively coarse grain size indicated high capacity equipment such as cones, spirals and jigs would be most suitable for cassiterite recovery, at least as a preliminary concentration stage to remove the bulk of the material processed as a barren tailings.

5) Define the performance of the circuit in terms of metal recovery and concentrate grade. This obviously involves the isolation and investigation of metal losses.

6) Estimate the concentrations of impurities in the concentrate and define processing to enable their rejection.

7) Combine the findings from above to define the optimum processing flowsheet, specify equipment requirements and costs and to estimate operating costs.

3.7 Samples Tested

In all, 11 samples from 9 diamond drill holes, and 4 bulk samples were tested. The samples were collected under geological supervision. Considerable consideration was given to the sites for their collection in order to ensure that they were representative of the orebody and contained typical mineralisation. The DDH samples were used for liberation studies. The majority of the metallurgical testwork was carried out on the four bulk samples. Figure 2 shows a plan of the deposit and the relative position of samples used.

3.7.1 Bulk Sample No. 1

Approximately three tonnes of sample were collected from the eastern face of the old workings. The sample was obtained by blasting and was of mineralised greisen granite which contained disseminated cassiterite and traces of sulfides. The sample assayed 0.35 per cent tin.

3.7.2 Bulk Sample No. 2

Approximately three tonnes of mineralised greisen granite were collected from the floor of the open pit close to the site of DDH BT71. The sample was collected by drilling 25m length, PQ holes and contained fine to medium grains of cassiterite erratically disseminated throughout the core. The sample assayed 0.29 per cent tin.

3.7.3 Bulk Sample No. 3A

Bulk sample No. 3 was collected by drilling PQ holes adjacent to DDH BT65 located on the hill north-east of the old workings. Due to differences in mineralogy the core was split into two subsamples, No. 3A and No. 3B.

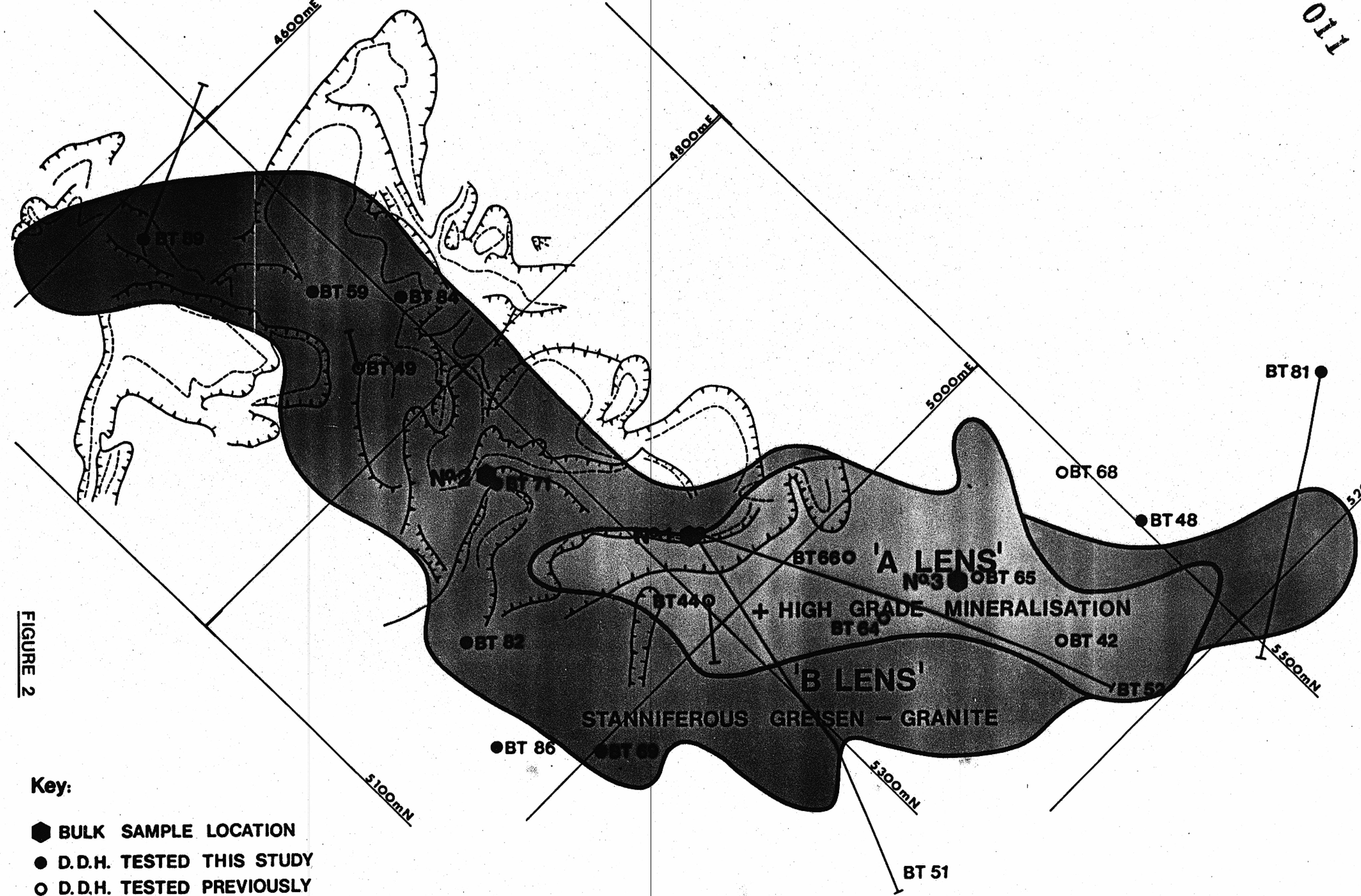

FIGURE 2

Key:
BULK SAMPLE LOCATION
D.D.H. TESTED THIS STUDY
D.D.H. TESTED PREVIOUSLY

ANCHOR TIN MINE LOCATION PLAN

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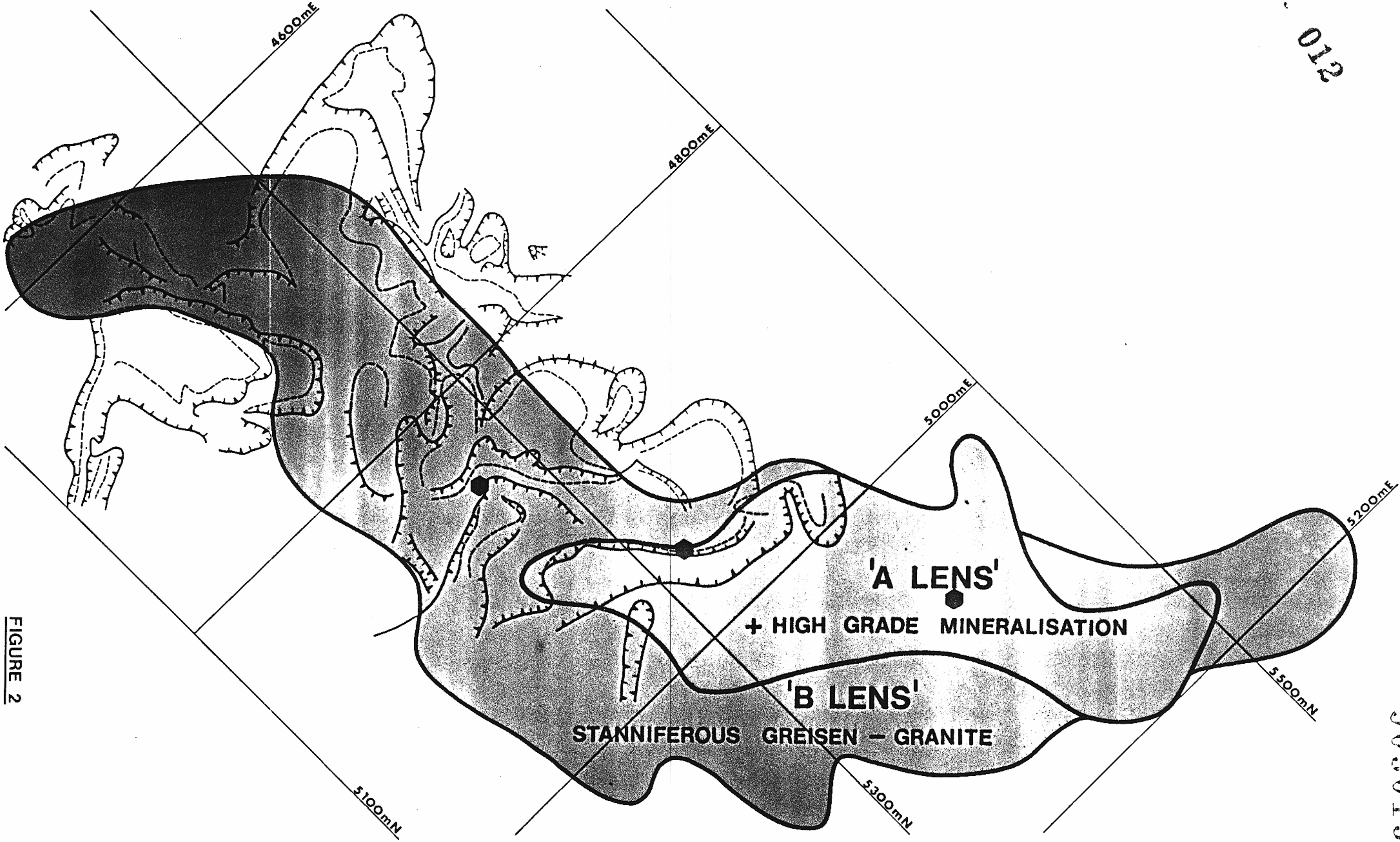

FIGURE 2

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ANCHOR TIN MINE LOCATION PLAN

Sample 3A consisted of coarse grained granular quartz-mica-topaz rock and minor greisen granite. Cassiterite was erratically disseminated as in other samples although comparatively more abundant and coarser grained. Copper and zinc sulfides were also more abundant than in other samples. The sample assayed 0.67 per cent tin.

3.7.4 Bulk Sample No. 3B

Sample No. 3B is typical of the low grade greisen granites obtained in other samples. The sample assayed 0.12 per cent tin.

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4. FLOWSHEET DEVELOPMENT

4.1 Summary of Testwork

4.1.1 Diamond Drill Core Samples

Samples from diamond drill holes were finely crushed and sized. Size fractions were separated using heavy liquids. The results from these tests enabled the required fineness of grind to be defined.

4.1.2 Bulk Samples

Subsamples from the four bulk samples were used for:

a) grindability tests to enable crushing and grinding equipment requirements to be defined.

b) heavy medium separation tests to determine if heavy medium separation was a feasible preconcentration stage.

c) heavy liquid testwork to compare liberation with that estimated from DDH samples and as an indicator to possible recovery processes.

d) gravity recovery testwork to enable the cassiterite recovery process to be defined and equipment requirements determined.

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4.2 Crushing

4.2.1 Introduction

Due to the fact that cassiterite grainsize is considerably finer than the size of run-of-mine rock it is necessary to reduce the rock to a size at which cassiterite is liberated from gangue material. Capital and operating costs are reduced and efficiency is increased if several stages rather than a single stage of size reduction are used. Again, due to cost savings and improved efficiency, crushing machines are used for size reduction of coarsest particles and grinding machines are used for the final reduction to the fine sizes required.

A further advantage of staged size reduction is that size reduction to relatively coarse particle sizes may liberate sufficient barren material to make a preconcentration process such as heavy medium separation feasible, thereby eliminating the unnecessary further treatment of valueless waste.

4.2.2 Crushing Requirements

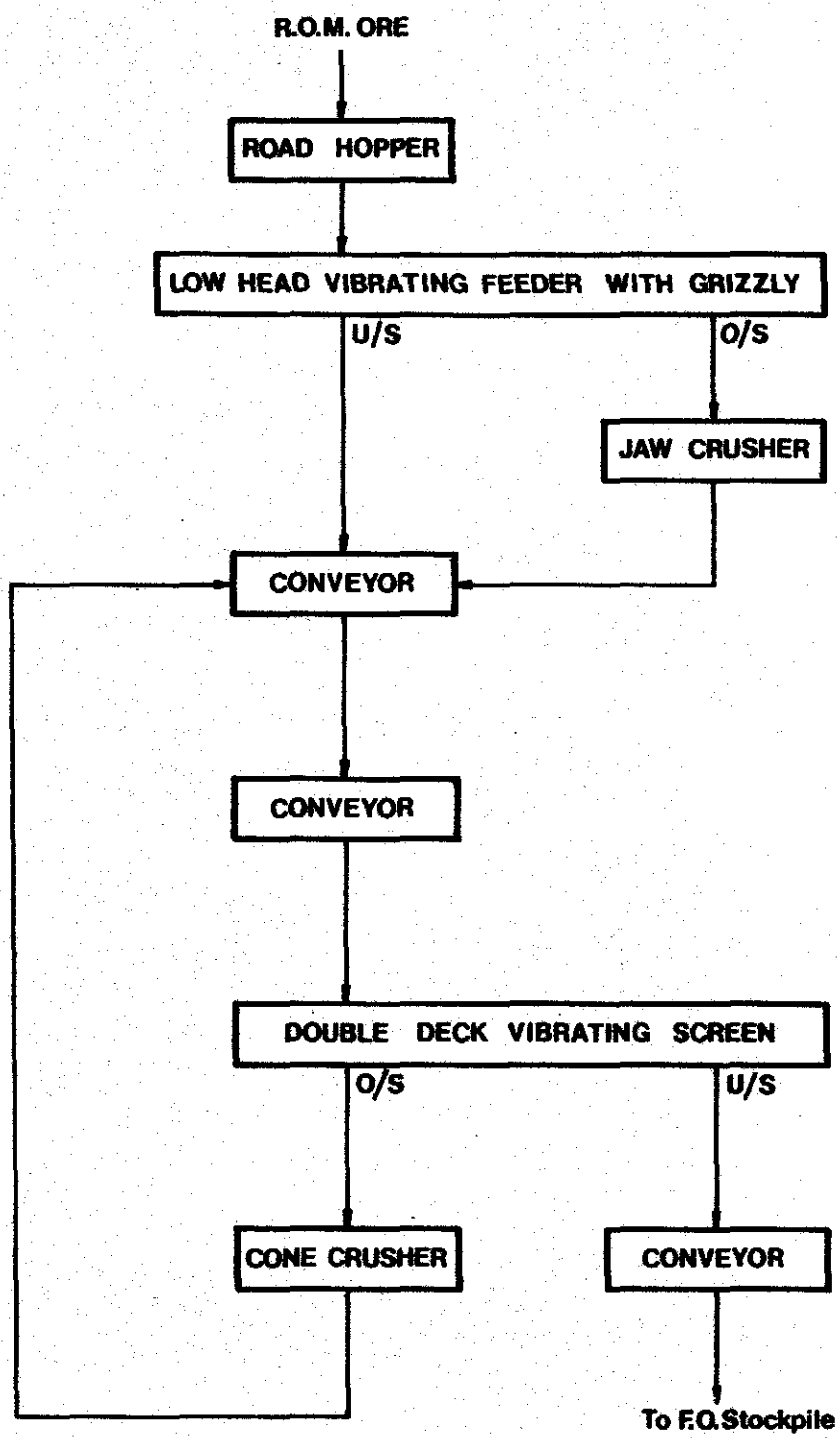
Costs are related to the size of machine, i.e. the bigger the machine, the more expensive. As costs are to be minimised, the smallest crushing machines should be selected. The minimum size of a crushing machine is limited by:

a) the size of the feed to the crusher.

b) the throughput or tonnage required.

The maximum size of ore from the mine has been estimated as being approximately 460mm (18 inches) (G. Northcote, personal communication, 10.11.82).

Product size can be defined as a size at which it becomes cheaper to reduce size further using grinding rather than crushing equipment.



BLUE TIER PROPOSED CRUSHING CIRCUIT

FIGURE 3

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4.2.3 Definition of the Circuit

Allis Chalmers were requested to assist and advise during the evaluation of the various options available to meet the requirements and constraints outlined above. Standard manufacturers tables were used for equipment selection.

4.2.4 The Circuit

The circuit proposed is outlined in Figure 3. Run-of-mine ore is dumped into a road hopper. The ore is removed from the hopper by a low head vibrating feeder which is fitted with grizzly bars to scalp fines from primary crusher feed. Primary crushing is performed in a 0.9m x 1.2m, single toggle jaw crusher. Jaw crusher discharge and grizzly undersize are combined and conveyed to a double deck screen. Minus 19mm fines from the screen are conveyed to a fine ore stockpile whilst screen oversize gravitates to the secondary crusher. Secondary crushing is carried out in a 1.5m diameter cone crusher and crusher product is passed back to the double deck screen.

This circuit has a capacity of 220 tonnes per hour and produces a product which is 80 per cent passing (or finer than) 14mm.

4.2.5 Discussion of Circuit

The circuit proposed was selected as the lowest cost option which would adequately perform the required function.

Three stage crushing was examined as this enabled smaller equipment of lower capital cost to be considered. The lower capital costs available using three stage crushing were offset by increases in conveyor, screen, etc. and installation costs.

The grizzly was included in the low head vibrating feeder as this offered significant reductions in wear of the crushing faces of the jaw crusher at a small additional cost.

A double toggle jaw crusher was seen to unnecessarily increase capital and installation costs.

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Closed circuit crushing was selected in order to maintain control of the circuit product size. This permits grinding circuit feed rate to be maximised and maintained and can also result in improved mill availability.

A surge bin between the screen and secondary crusher was considered but the operating advantages were outweighed by the costs of the bin.

4.2.6 Equipment Details and Costs

Details of the equipment included in the circuit are tabulated along with estimates of the costs of these items in Table 1.

TABLE 1 CRUSHING CIRCUIT - EQUIPMENT DETAILS AND COSTS

ITEM	DETAILS	Total kW	Budget $	Install $	Total $
1 Crushers					
Primary - Jaw	0.9m x 1.2m, single toggle	130	125,000	60,000	185,000
Secondary - Cone	7-60 Hydrocone, 1.5m diam. 178mm (medium) chamber	225	265,000	80,000	345,000
2 Screen	2.4 x 6.1m Ripl-Flo, double deck, 50mm and 19mm aperture	30	46,000	50,000	96,000
3 Feeder	1.2 x 3.7m Low Head Vibrating (with anti-current coupling forV.S. drive) with 1.2m of grizzly	15	39,000	20,000	59,000
4 Road Hopper	100 tonne capacity - $48m^3$ - steel				70,000
5 Conveyors					
Crusher Discharge	61m with 42m at 18° incline	30			100,000
Screen Feed	44m at 18° incline	30			70,000
Fine Ore Stockpile Feed	30m	30			50,000
Transfer Tower					70,000
6 Miscellaneous					
V-belt drives	All crushers				5,000
Motors	All crushers and screen				70,000
Dust Collection/ Ducting/Pump					70,000
Overhead Crane	1 x 20t capacity				60,000
Metal Protection	1 magnet located at discharge onto screen				30,000
7 Crusher Building	12m x 20m x 26m high				360,000
8 Electrical	10% of equipment and building costs				170,000
					1,810,000

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4.3 Heavy Medium Separation

Heavy medium separation is commonly used to preconcentrate minerals. Any ore which exhibits sufficient liberation to enable a separation of particles (due to differences in specific gravity) into streams containing particles which will repay the cost of further treatment and those which will not is amenable to this process.

Samples No. 1 and No. 2 were submitted to AMDEL for Heavy Medium Separation Evaluation. Results are detailed in Appendix I. The results indicate the material is not suitable for preconcentration using heavy medium separation. No further testwork was carried out.

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020 4.4 Grinding

4.4.1 Introduction

Grinding is required to comminute ore from a size below which it is uneconomical and inefficient to use crushing machines to a size at which liberation of valuable minerals is achieved, thus permitting the efficient recovery of valuable minerals.

4.4.2 Grinding Requirements

As for crushing machines, costs increase as the size of equipment increases. The minimum size of grinding mill which can be used is governed by the amount of energy required to reduce the ore from feed size to the required product size at the required treatment rate. Defining feed size, required product size, work index (a measure of the amount of energy required to reduce the ore from a standard feed size to a standard product size in a standard mill) and required throughput enables mill size to be defined.

4.4.3 Feed Size

The feed size from the mill is set as the product size from the crusher. In this case it is 80 per cent passing 14 mm.

4.4.4 Product Size

Heavy liquid testwork results from the eleven DDH samples and the four bulk samples are detailed in Appendix II. These results indicate a product which is 80 per cent passing 350 μm is required.

021 4.4.5 Work Index

Data from grindability tests is shown in Table 2

Table 2 Grindability Results

	Sample No		
	1	2	3A
Rod Mill work index at 1700µm (kWh/t)	11.2	11.0	10.1
Ball mill work index at 600µm (kWh/t)	14.0	13.0	12.9
Ball mill work index at 300µm (kWh/t)	13.2	13.1	13.2
Impact crushing work index	10.8		
Abrasion Index	0.2095		
Sp.gr.	2.8	2.6	2.8

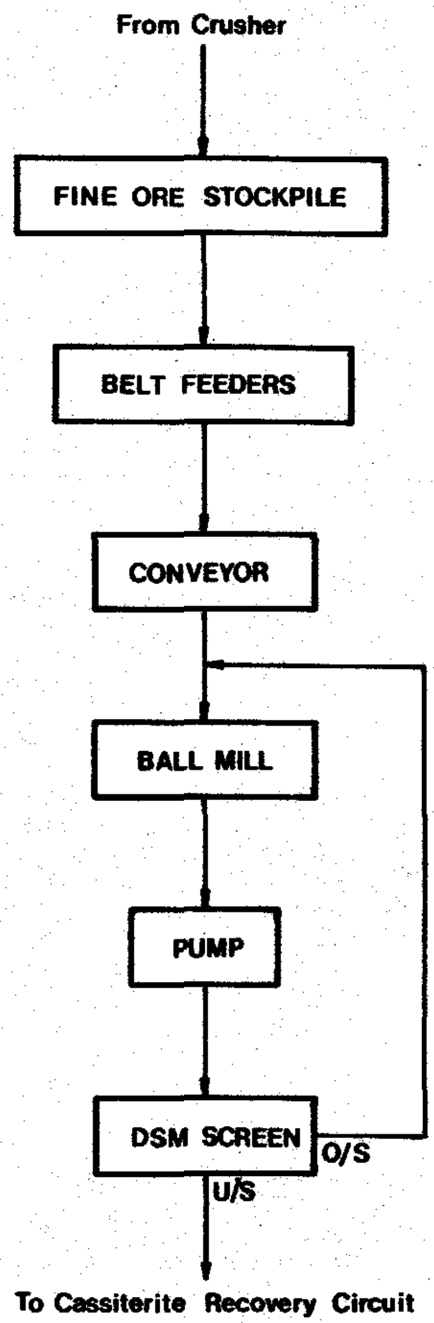
In all calculations the highest values were used rather than an average so that, if anything, actual capacity will be slightly higher than the estimated capacity. It was felt that to inadvertantly undersize grinding equipment could have a serious effect on a marginal operation.

4.4.6 Capacity

In order to define the hourly treatment rate of the mill it is first necessary to define operating hours and annual treatment rate.

An examination of a five day operating week versus a seven day operating week was made. Operating seven days per week enables smaller capacity equipment to be utilised with a corresponding reduction in capital cost. However labour costs increase as the number of employees must be increased and penalty rates must be paid for weekend work. Calculations indicated that for a ten year mine life, capital costs must be reduced by approximately 1,000,000 dollars in order to offset increased labour costs and therefore justify extending operating time to seven days per week. A reduction in capital costs of 1,000,000 dollars proved impractical due to the scale of this operation.

It was therefore decided to assume that the mill would operate for 51 weeks per year, 5 days per week and 24 hours per day. It was assumed that the mill would operate for 95 percent of available time. This gives a total of 5814 operating hours per year.



GRINDING CIRCUIT FLOW SHEET

FIGURE 4

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A base case in which the annual treatment rate of the mill is 350,000 tonnes per year was examined.

This enabled the required mill capacity to be defined as 61 tonnes per hour.

4.4.7 Definition of the Circuit

The circuit was defined from basic grinding equations and calculations. Allis Chalmers were called upon to study and make recommendations upon the proposed circuit.

4.4.8 The Circuit

The proposed circuit is outlined in Figure 4. The circuit consists of a fine ore stockpile to provide storage capacity after crushing. Ore from the fine ore stockpile is conveyed to a 3.35m by 3.66m diameter ball mill. Ball mill discharge is pumped to D.S.M. screens (1mm aperture) and screen oversize is returned to the ball mill. Screen undersize passes to the gravity concentration section.

The ball mill used in the circuit is an overflow discharge type, lined with rubber and charged with balls with a maximum size of 80mm.

4.4.9 Discussion of the Circuit

The circuit proposed was selected as the lowest cost option that would enable the reduction of crusher product to the required size at the required throughput.

Ore storage for four operating shifts (2,000 tonnes) was thought necessary. Fine ore bins were examined and rejected due to prohibitive costs. Stockpiling followed by underground reclamation was selected. A covered stockpile was rejected, again to minimise unnecessary costs (although the site has an annual rainfall of approximately 1,500mm, Rossarden, which has a similar climate, operated with open stockpiles and reported no major problems).

Single stage grinding was preferred to multi-stage grinding due to reductions in costs.

Rod milling was considered as an alternative to ball milling. The relatively high reduction ratios involved reduce the efficiency of rod milling and would therefore necessitate an increase in the number of grinding mills required.

An overflow discharge type mill was selected over a grate discharge type as lower operatings costs and improvements in efficiency and availability are possible.

Screens rather than hydrocyclones were selected for classification duty. This is because cassiterite, which is of a relatively high specific gravity, misreports to hydrocyclone underflow resulting in overgrinding which in turn leads to poorer mineral recovery. D.S.M. screens are preferred to wedge wire screens due to improved classification efficiency, reductions in wear costs and reductions in floor area requirements per unit capacity. Rubber mill liners are recommended due to reduced wear and noise levels and ease of replacement. The maximum ball size is governed by the size of feed and the mill size.

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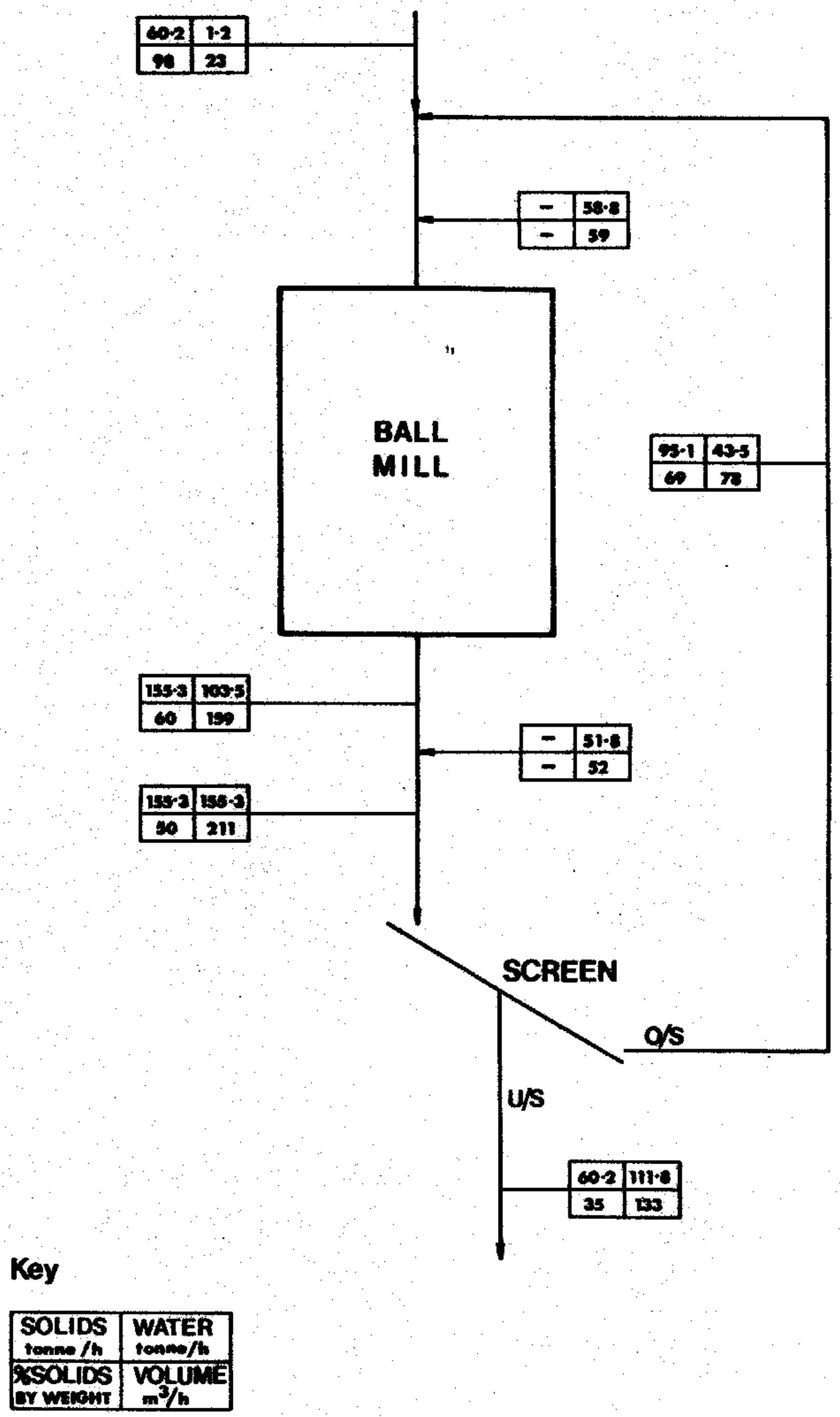

GRINDING CIRCUIT MASS BALANCE

FIGURE 5

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4.4.10 Equipment Details and Costs

Details of the equipment included in the grinding circuit are tabulated along with cost estimates in Table 3.

TABLE 3 GRINDING CIRCUIT - EQUIPMENT DETAILS AND COSTS

ITEM		Total kW	Budget $	Install $	Total $
1 Ball Mill	3.35m x 3.66m, rubber lined	570	610,000	160,000	770,000
2 Fine Ore Feeding					
Belt Feeders	2 off V/S belts	10			30,000
Conveyor	30m with 13m at 18° incline	11			40,000
3 DSM Screen	DSM panels, 1mm aperture, allow 8 x 0.9m panels				110,000
4 Pumps	TPH Solids % Solids m^3/hr Pump Size				
Ball Mill Discharge	155 50 211 6/4	56			37,000
Spillage		10			10,000
5 Ball Mill Charge	18.3t of 80mm, 23.1t of 65mm, 9.1t of 50mm, 3.2t of 25mm		37,000		37,000
					1,034,000

Figure 5 shows an expected mass balance for the circuit (operating with a 250 per cent circulating load.)

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4.5 Cassiterite Recovery

4.5.1 Introduction

Minerals are beneficiated by exploiting the differences in physical and chemical properties of the economic and gangue minerals. In processing cassiterite bearing ores, gravity separation is usually used where cassiterite grain size is sufficiently coarse to utilise such a process efficiently. Gravity separation processes exploit the relatively high specific gravity (sp.gr. = 7.0) of cassiterite to produce a separation and upgrading from less dense gangue material (sp.gr. <3.0).

Where cassiterite grain size is too fine to enable the efficient use of gravity separation processes, froth flotation is commonly used to effect concentration. Froth flotation exploits the chemical properties of cassiterite.

4.5.2 Testwork

The heavy liquid separation tests carried out indicated gravity separation to be a suitable technique for the recovery of cassiterite from Blue Tier mineralisation. The relatively coarse grain size and high degree of liberation, combined with the low grade of the deposit, suggested that machines of high capacity per unit floor area should be used to reject the bulk of the ground material to a barren tailing stream whilst producing an upgraded concentrate. Consequently testwork was conducted to determine the suitability of cones, spirals and jigs as concentrating stages. Cone and spiral testwork was initially carried out on a subsample of bulk sample No. 1 by Mineral Deposits. Jigging testwork on bulk sample No. 1 was carried out by Renison personnel concurrent with the Mineral Deposit testwork. Although jigging results were good the problems experienced during the testwork indicated that cones and spirals may have advantages over jigs in ease and simplicity. Cones and spirals were also expected to show lower maintenance costs than jigs.

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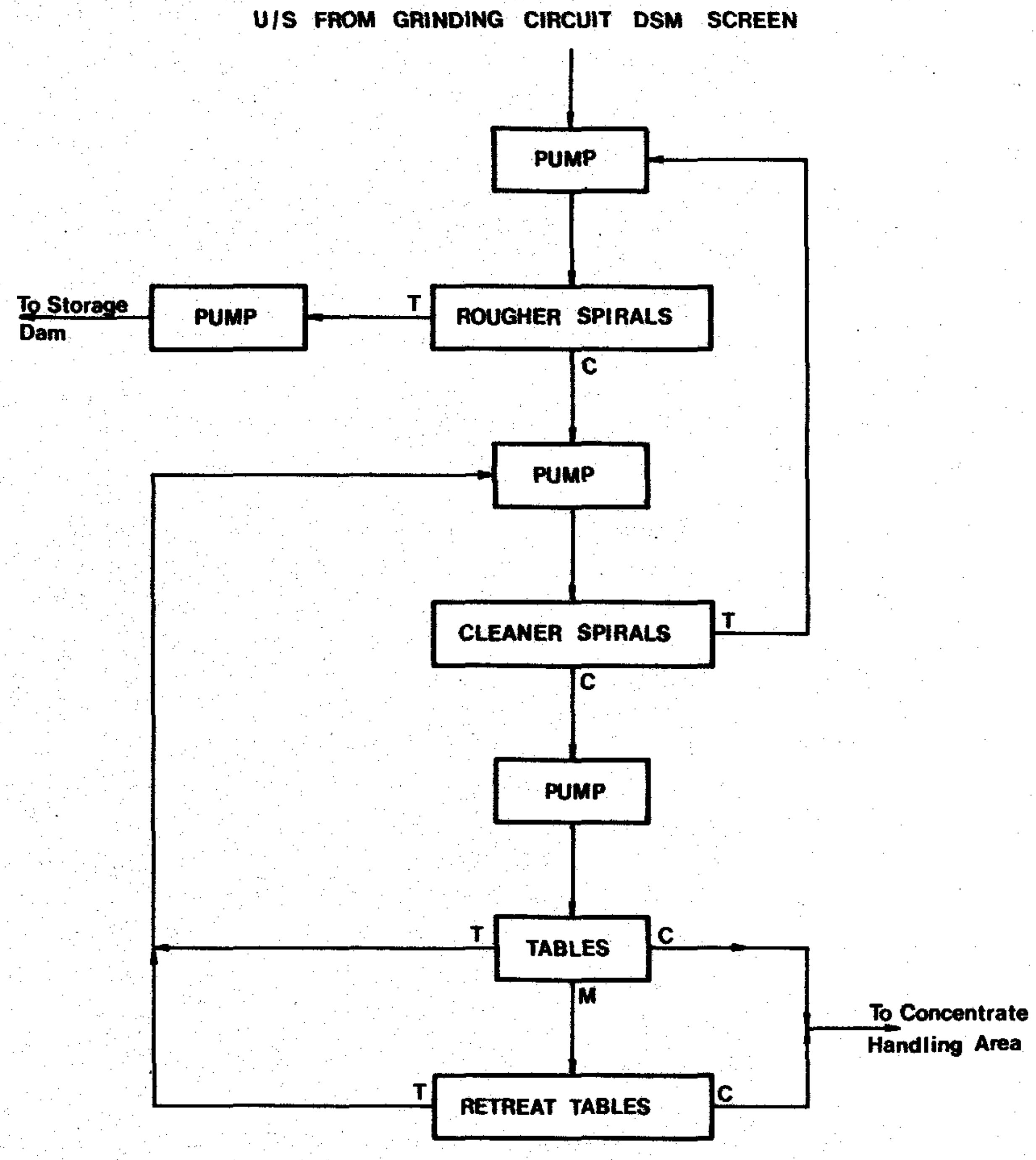

CASSITERITE RECOVERY FLOW SHEET

FIGURE 6

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Upon the completion of the testwork conducted by Mineral Deposits sufficiently encouraging results were obtained to justify the cancellation of further jigging testwork which had been planned. Instead, additional spiral testwork was carried out by Renison personnel on bulk samples No. 2 and No. 3B at the Mines Department in Launceston. Details of gravity separation testwork are given in Appendix III.

4.5.3 Circuit Requirements

The circuit was required to provide maximum metal recovery to a concentrate of maximum grade at the required treatment rate. It was also required that the circuit be simple and easy to operate thus minimising capital and operating costs.

4.5.4 Definition of the Circuit

The circuit was selected following interpretation of gravity testwork results. Scale-up from laboratory testing to full scale operation is a relatively simple matter for gravity separation processes. Simple simulation techniques were used to predict circuit performance and define equipment capacities. Mineral Deposits assisted with the interpretation of testwork results and the derivation of the proposed circuit.

4.5.5 The Circuit

Figure 6 outlines the proposed circuit. Screen undersize from the grinding circuit is pumped to roughing spirals where a barren tailing is rejected. Concentrate from the spiral is further upgraded by cleaning spirals. The cleaner spiral tailing is recycled back to the roughing spirals. Cleaner spiral concentrate is further upgraded on shaking tables. Table tailing is recycled to the cleaner spirals. Table concentrate passes to the concentrate handling area. Table middling is retreated on further tables, the concentrate from which also passes to the concentrate handling area whilst the tailing and middling are returned to the cleaner spiral.

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4.5.6 Discussion of the Circuit

As mentioned in 4.5.2 above, cones and spirals appeared to have significant advantages over jigs. In addition to the points mentioned previously, there has been a considerable increase in interest in recent years in cone and spiral technology with corresponding improvements in design, performance and manufacturer interest. Consequently high efficiency, custom made plant packages are available which has led to reductions in capital and installation costs. For these reasons cones and spirals rather than jigs were considered suitable for the circuit.

Gravity testwork indicated that cones and spirals produced similar results. Cones are generally lower capital cost items than spirals however they require additional costs to provide dewatering of feed material (as they operate at higher pulp densities than spirals), and increased head room. Cones are also limited by the minimum tonnage they can be used to process. It is for these reasons that spirals were selected for both roughing and cleaning duties.

A single stage of concentrate cleaning using shaking tables was initially envisaged. The two stage table circuit proposed was shown to produce significantly higher concentrate grades with a minor reduction in metal recovery and at a modest cost.

Desliming grinding circuit product prior to the cassiterite recovery circuit was considered. Testwork showed that desliming was not only unnecessary but also resulted in an unnecessary loss of mineral. The major losses from the circuit are in the finest size fractions (finer than 50μm), where the efficiency of gravity separation decreases, and the coarsest size fractions (coarser than 200μm) where cassiterite liberation is incomplete. Froth flotation could conceivably be employed to recover a portion of the fine cassiterite lost from the proposed circuit, whilst regrinding the coarse fraction, should improve liberation and consequently mineral recovery. In keeping with the philosophy of minimising capital and operating costs neither a froth flotation circuit nor a regrind mill were considered for inclusion in the proposed circuit. They may however prove to be viable options for increasing revenue during the operating life of the mine.

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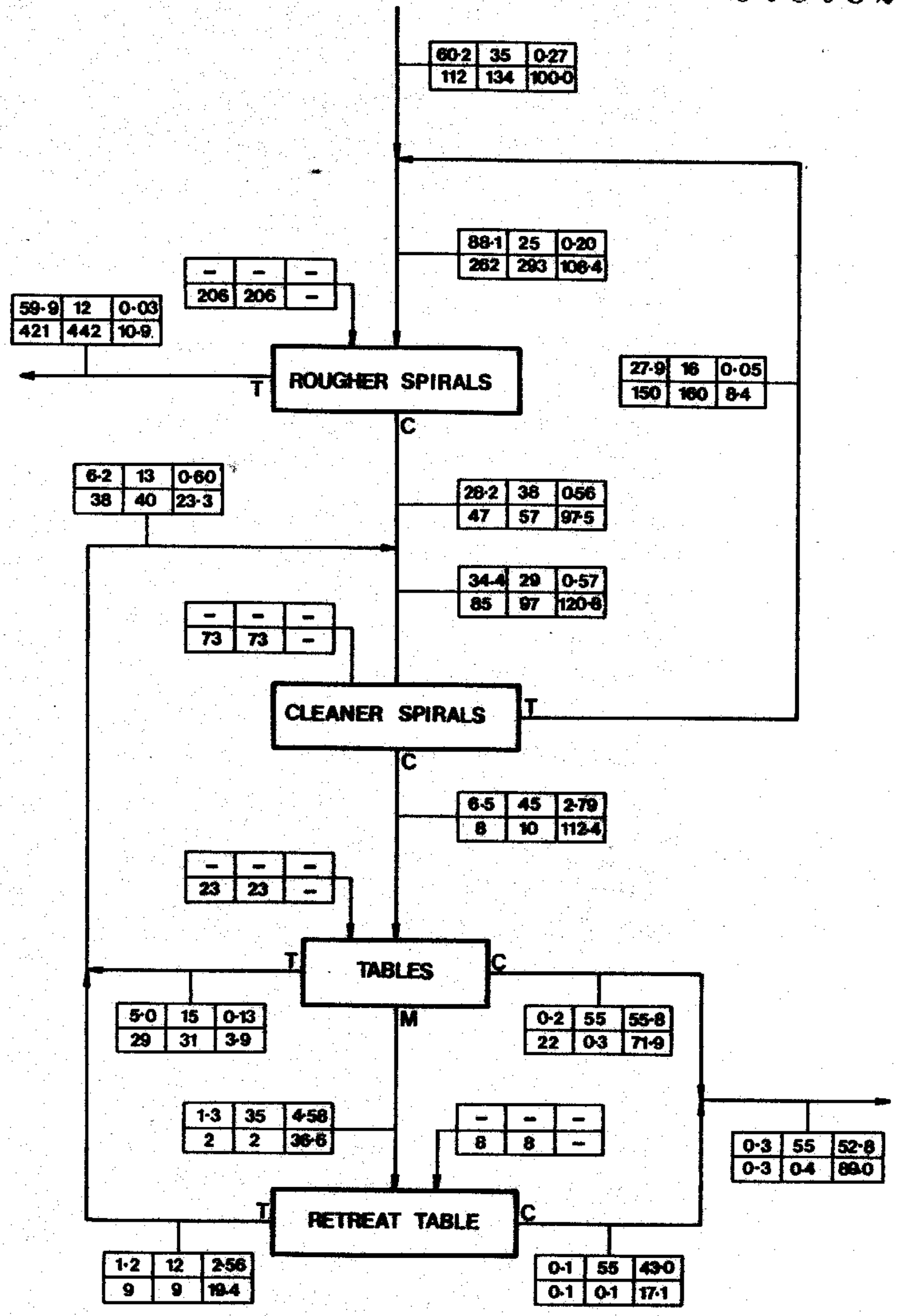

Key :

SOLIDS tonne/h	%SOLIDS BY WEIGHT	ASSAY % Sn
WATER tonne/h	VOLUME m³/h	% TIN DISTRIBUTION

GRAVITY CIRCUIT MASS BALANCE

FIGURE 7

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4.5.7 Equipment Details and Costs

Table 4 details the equipment requirements of the circuit along with estimates of equipment costs.

TABLE 4 CASSITERITE RECOVERY - EQUIPMENT DETAILS AND COSTS

ITEM						Total kW	Budget $	Install $	Total $
1 Spirals									
Rougher	38 off twin start, 7 turn spirals						114,000	35,000	149,000
Cleaner	12 off twin start, 7 turn spirals						36,000	15,000	51,000
2 Tables	8 off Holman Tables					12	96,000	70,000	166,000
3 Pumps	TPH Solids	% Solids	m^3/hr	Pump Size					
Rougher Feed	88	25	293	8/6		37			46,000
Mill Tailings	60	12	442	10/8		56			60,000
Cleaner Feed	35	29	97	6/4		22			32,000
Table Feed	7	45	10	2/1½		4			13,000
Spillage						10			10,000
									527,000

Figure 7 shows a mass balance for the circuit under normal operating conditions when treating material of a similar grade to that of the reserve.

4.6 Concentrate Handling

4.6.1 Introduction

Any concentrates produced by a mineral separation process requires some further processing. This can range from dewatering and packaging through to further upgrading processes.

4.6.2 Circuit Requirements

The concentrate handling circuit is required to produce a concentrate of saleable grade, packaged in a manner suitable for transport to a smelter, with minimum loss of metal. It is also required that the circuit permits accurate accounting of production.

4.6.3 The Circuit

Concentrates from the cassiterite recovery circuit gravitate into 44 gallon drums. Water overflows from the drums and is reintroduced into the cassiterite recovery circuit via a spillage pump. The drums are decanted, topped up, sampled, weighed and sealed on a daily basis. Approximately six tonnes of concentrate are handled by this circuit each weekday.

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4.6.4 Concentrate Analysis

Predicted analysis figures for the concentrates produced are shown along with ranges of assays expected in Table 5.

Table 5 Predicted Concentrate Analysis

Element	Assay	Range
Sn	52.8%	50-55%
Fe	3.6%	1.2-5.9%
S	1.5%	<0.1-3%
As	<0.01%	*
Cu	0.6%	0.03-1.2%
Pb	0.10%	0.07-0.12%
Zn	1.1%	0.08-2.1%
Ag	86 ppm	5 - 166 ppm
Mo	0.20%	0.14-0.26%
Bi	0.22%	0.09-0.35%
WO_3	1.0%	0.4-1.5%

* no assays reported exceeded 0.01 per cent

An examination of base metal assays for the DDH and bulk samples showed the deposit to consist of approximately equal amounts of material containing high and low base metal levels. Predicted assays were derived by proportionally combining results from the bulk samples which contained high and low levels of base metals.

The ranges were produced by taking the highest and lowest assays produced.

4.6.5 Discussion of the Circuit

The major diluent in the concentrate is topaz. Base metal sulfides and biotite also contaminate the concentrate.

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Further processing to upgrade the concentrate by removing topaz or the base metal sulfides was considered but is not recommended as:

1) topaz has a relatively high specific gravity (sp.gr. = 3.5) which would reduce the efficiency of any separation.

2) levels of base metals, although predicted to be significant at times, would in general only enable an upgrading of the concentrate by four per cent tin.

3) the low tonnage of concentrates and low levels of base metal sulfides in combination with the complexity of any sulfide concentrate produced would indicate the base metal sulfides would be difficult to market thereby eliminating any potential economic value they may be seen to have.

4) any further upgrading process would reduce the overall recovery of tin metal.

Further processing to reduce the levels of penalty elements in the concentrate was considered. Predicted concentrates shown in Table 5 contain relatively high quantities of base metals, specifically copper, zinc and bismuth. Such concentrates are marketable (R. Boyer, personal communication, 3.2.83) however concentrates produced containing the upper limits of the ranges shown in Table 5 would be approaching the limits accepted by smelters. Further processing could be employed to remove base metal impurities thereby minimising smelting penalties. However blending of concentrate lots would overcome this problem should it arise.

It is recommended that, during any future definitive investigation of the project, typical contract terms should be defined and that the effect of contaminent levels on revenue should be estimated.

Concentrate drying was considered unnecessary.

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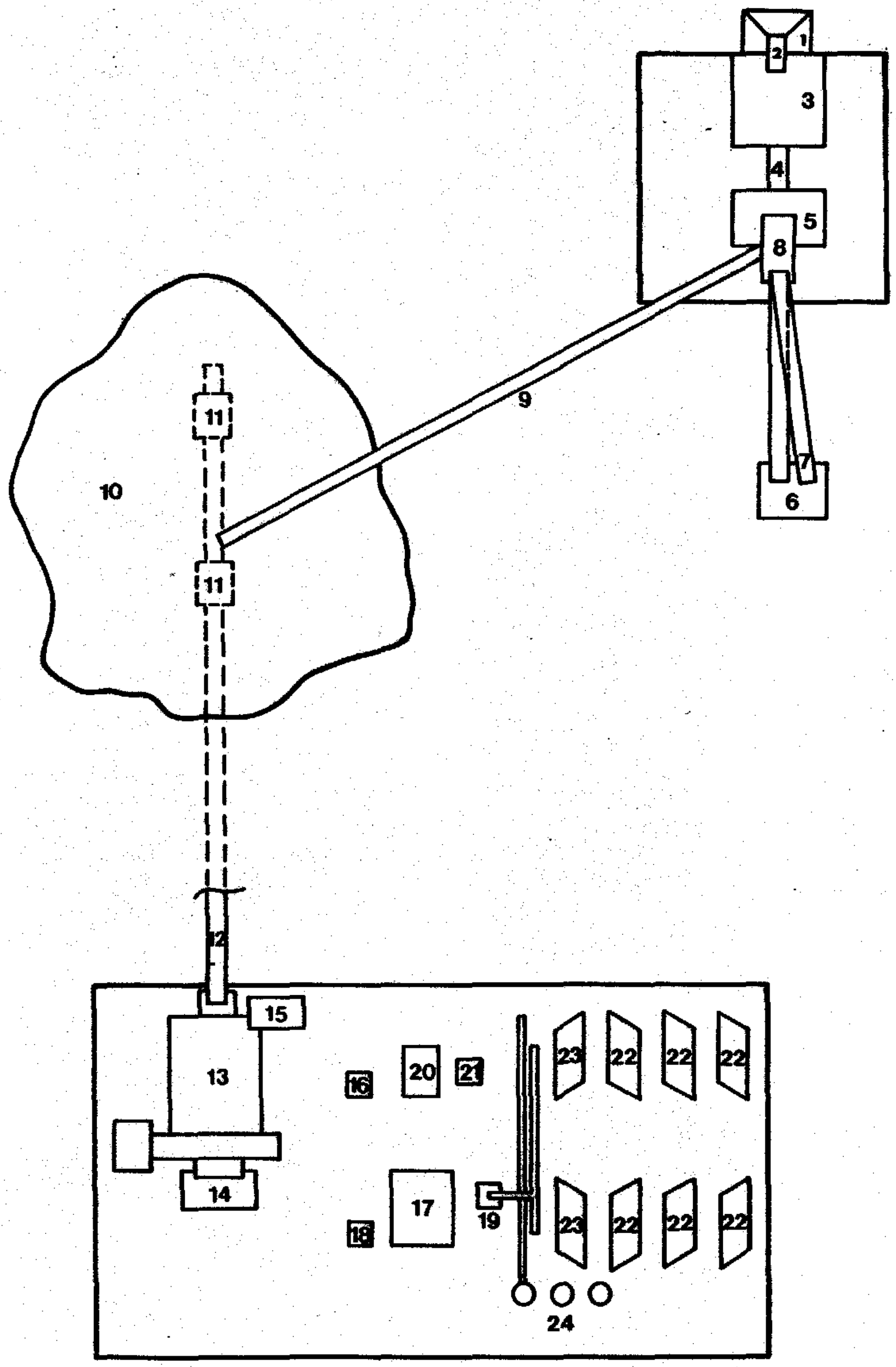
The concentrate handling section is labour intensive. Personnel utilised in this area also perform other necessary duties in other areas of the plant. Any mechanisation of this area would be difficult and costly without significantly reducing labour requirements.

4.6.6 Equipment Details and Costs

Table 6 details the equipment requirements of the section along with estimates of costs. Other miscellaneous equipment and mill buildings are also shown in Table 6.

TABLE 6 CONCENTRATE HANDLING AND GENERAL EQUIPMENT DETAILS AND COSTS

ITEM	DETAILS	Total kW	Budget $	Install $	Total $
1 Scales					5,000
2 Crane	1 x 5t capacity				20,000
3 Miscellaneous Equipment					
a) Water Reticulation					
i) Tank	$250m^3$ i.e. 30 minutes capacity - 4m x 9m ∅				25,000
ii) Constant Head Tank	$30m^3$ i.e. 1.72m x 1.72m ∅				4,000
iii) Pump	$430m^3$/hr	56			20,000
b) Assaying/Sampling					
i) Auto samplers	For feed and tailings				10,000
ii) Analyser	AMDEL Mineral Analyser		16,000		16,000
iii) Sample preparation					10,000
c) Mill Building	18m x 30m x 10m high				600,000
d) General					
i) Distributors	for spirals and tables		10,000	10,000	20,000
ii) Weightometer	for ball mill feed				20,000
iii) Crane	for grinding area - 1 x 10 tonne unit				35,000
e) Electrical	22% of plant and building				500,000
f) Piping	27% of plant only				460,000
g) Instruments	10% of plant only				170,000
					1,915,000



PROPOSED LAYOUT

NOTE: Not to scale. (For details see table 7)

FIGURE 8

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4.7 Miscellaneous

4.7.1 Layout

A layout similar to that outlined in Figure 8 is proposed. Equiment shown in Figure 8 is tabulated below.

Table 7 Key to Figure 8

1	Road Hopper
2	Low Head Feeder
3	Jaw Crusher
4	Crusher Discharge Conveyor
5	Cone Crusher
6	Transfer Tower
7	Screen Feed Conveyor
8	Double Deck Screen
9	Fine Ore Stockpile Feed Conveyor
10	Fine Ore Stockpile
11	Belt Feeders
12	Mill Feed Conveyor
13	Ball Mill
14	Ball Mill Discharge Pump
15	DSM Screen
16	Rougher Spiral Feed Pump
17	Rougher Spirals
18	Mill Tailing Pump
19	Cleaner Spiral Feed Pump
20	Cleaner Spirals
21	Table Feed Pump
22	Tables
23	Retreat Tables
24	Concentrate Drums

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In the proposed layout the table floor is inclined. This enables table products to be gravitated minimising pump requirements. Excavating a pump floor or suspending an operating floor to enable tables to be located at the same level would be more costly than producing an inclined floor. Of course, costs can be minimised by exploiting any natural stope.

4.7.2 Pumps

Spillage pumps have been provided in the grinding and cassiterite recovery/concentrate handling sections. Standby pumps were not considered for the main pumping duties due to the small scale and relatively low running time of the operation.

The assumption that the mill with no standby pumps would run for 95 per cent of operational time is considered optimistic. The impact on equipment requirements of overestimating running time can be shown to be negligible.

4.7.3 Water

Approximately four hundred and twenty cubic metres per hour of water are required to operate the concentrator and a tank with 30 minutes capacity has been included. Water from this tank is pumped to a constant head tank located in the roof of the concentrator. A suitable supply of water has been assumed.

4.7.4 Tailing Disposal

As with the I.F.S., tailing disposal is one of the most difficult problems of the deposit. It is beyond the scope of this report to select storage locations and dam building methods. It should be noted that if an earthen retaining wall is not used some form of thickening of tailing will have to be employed. In addition, extra labour will be required for the construction of the dam walls.

Regardless of the approach to tailing storage, some form of water reclamation is advised.

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4.7.5 Sampling/Assaying

Samples of mill feed and tailing should be collected for accounting purposes.

Automatic samplers were included in the proposed circuit in an attempt to maximise sample reliability.

Full assay facilities are difficult to justify for the small number of concentrate, accounting and control samples.

A portable isotope X-ray analyser is considered sufficient for control samples.

It is recommended that accounting and concentrate samples should be processed at an off-site facility.

An allowance of ten thousand dollars has been included in the capital costs for miscellaneous sample preparation facilities.

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5. PROJECT COSTS

5.1 Introduction

The flowsheet developed in the previous section assumed an annual throughput of 350,000 tonnes for a life of ten years. Capital and operating cost estimates are detailed below for this base case.

In the base case, operations are carried out for five of the seven days in a week. A second case, (Case A) in which operations are extended to seven days per week thus increasing annual throughput to 500,000 tonnes, is also examined.

In both of these cases the capacity of the crusher is far in excess of that of downstream equipment. A third case (Case B) has been considered in which equipment downstream of the crusher is expanded in order to fully utilise crusher capacity. In this case annual throughput is 1,100,000 tonnes of ore.

5.2 Base Case - 350,000 Tonnes per Year

5.2.1 Capital Costs

Capital cost estimates are shown in Table 7

Table 7 - 350,000 tpa Capital Costs

Section	$
Crushing (Table 1, 4.2.6)	1,810,000
Grinding (Table 3, 4.4.10)	1,034,000
Cassiterite Recovery (Table 4, 4.5.7)	527,000
Concentrate Handling and Miscellaneous (Table 6, 4.6.6)	1,915,000
Contingency (20% of total of above items)	1,057,000
Total	6,343,000

5.2.2 Operating Costs

5.2.2.1 Operating Labour

5.2.2.1.1 Manning

At the required annual throughput the crusher would have to be available for one shift per day, five days per week, 51 ~~days~~ weeks per year. The crusher would have to be running 80 per cent of this available time. One operator would be required with assistance from day services personnel for clean-up and operator training.

The grinding and cassiterite recovery sections are manned 24 hours per day, 5 days per week, 51 weeks per year. Three shifts of two men (one for each section) are therefore required. The work load of each man is low but the number of operators per shift should be maintained at two because, firstly, the grinding section operator will require assistance if he encounters problems with fine ore supply to the mill and secondly, conditions for security and safety are improved.

Four personnel are required as day crew for concentrate handling operations, clean-up duties and operator relief and training. Manning levels for the three cases are summarised in Table 8.

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TABLE 8 MANNING LEVELS

	Base Case	Case A	Case B
Crushing	1	1	4 x 2
Grinding	3 x 1	4 x 1	4 x 1
Cassiterite Recovery	3 x 1	4 x 1	4 x 1
Day Crew	4	5	8
Foremen	-	-	1
Metallurgist	1	1	1
Total	12	15	26

5.2.2.1.2 Costs

A base rate of \$6.60 per man hour (R. Cohen, personal communication, 2.2.83) has been assumed. An increment of \$6.67 per shift has been assumed for night and afternoon shift work. Overtime has been assumed to be worked at the rate of one overtime shift per fortnight per employee paid at double time. Labour overheads have been assumed as thirty per cent of total labour costs.

5.2.2.2 Operating Supervision

Shift supervision cannot be justified.

The entire metallurgical operation will be supervised by a metallurgist. A salary of \$30,000 (plus 30 per cent of this sum as overheads) has been assumed.

Other technical/support positions such as analyst or metallurgical accountant cannot be justified.

5.2.2.3 Operating Consumables

5.2.2.3.1 Grinding Media

Ball consumption was calculated using the abrasion index shown in Table 2 of section 4.4.5. A consumption of 0.94 kilograms of balls per tonne milled is estimated. Media costs were assumed to be 70 cents per kilogram.

5.2.2.3.2 Concentrate Handling

An annual cost of $14,000 for drums for concentrate despatch was estimated.

5.2.2.3.3 Administration

A cost of one cent per tonne milled was assumed for administration consumables. Assaying costs are estimated at 11 cents per tonne milled.

5.2.2.4 Power

Power costs have been estimated by totalling the power ratings of all motors in the proposed plant. 85 per cent usage has been assumed and this power rating has been multiplied by running hours to give a total kilowatt hour figure for the year. A power cost of six cents per kilowatt hour has been assumed (I. Wood, personal communication, 3.2.83).

5.2.2.5 Maintenance

For the crushing, grinding and cassiterite recovery sections maintenance costs of 38, 20 and 5 cents per tonne milled respectively have been estimated. For low maintenance areas such as concentrate handling and administration, costs of one cent per tonne milled have been assumed. Maintenance costs were based on data available from Renison. A 30 per cent loading for overheads has been assumed for maintenance labour.

5.2.2.6 Total Operating Costs

Operating costs are detailed in Table 9. A total operating cost of $3.16 per tonne milled is estimated.

TABLE 9 BASE CASE (350,000 tpa) - OPERATING COSTS

	Operating Labour			Operating Supervision			Operating Consumables			Power			Maintenance			Total		
	$	$/t	%	$	$/t	%	$	$/t	%	$	$/t	%	$	$/t	%	$	$/t	%
Crushing	16210	0.05	1.5	-	-	-	-	-	-	40784	0.12	3.7	133000	0.38	12.1	189994	0.55	17.3
Grinding	52030	0.15	4.7	-	-	-	230300	0.66	20.9	194810	0.56	17.7	70000	0.20	6.4	547140	1.57	49.7
Cassiterite Recovery	52030	0.15	4.7	-	-	-	-	-	-	44009	0.13	4.0	17500	0.05	1.6	113539	0.33	10.3
Concentrate Handling	64838	0.19	5.9	-	-	-	14000	0.04	1.3	3500	0.01	0.3	3500	0.01	0.3	85838	0.25	7.8
Administration + Overheads	55532	0.16	5.0	39000	0.11	3.5	43500	0.12	4.1	3500	0.01	0.3	21896	0.06	2.0	163428	0.46	14.9
Total	240640	0.70	21.8	39000	0.11	3.5	287800	0.82	23.4	286603	0.83	26.0	245896	0.70	22.4	1099939	3.16	100.0

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5.3 Case A - 500,000 Tonnes per Year

5.3.1 Capital Costs

As operating time is extended to seven days per week, continuous operation, standby pumps are justified to maintain levels of operating time. This increase in capital costs is shown in Table 10.

Table 10 - 500,000 tpa Capital Costs

SECTION	$
Crushing	1,810,000
Grinding	1,071,000
Cassiterite Recovery	678,000
Concentrate Handling and Miscellaneous	2,055,000
Contingency (20% of total of above items)	1,123,000
Total	6,737,000

5.3.2 Operating Costs

5.3.2.1 Operating Labour

5.3.2.1.1 Manning

As continuous seven day per week operation of the mill is required, crushing operation also has to be extended to seven eight hour shifts per week. No increase in manning levels for the crusher are required as the additional shifts required would be rostered as overtime shifts.

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Shift operators are increased by two to allow for continuous operation.

Day crew personnel are increased by one to allow for the increased quantity of concentrates handled (See Table 8).

5.3.2.1.2 Costs

Costs are calculated as in previous sections. Saturday and Sunday hourly rates are at time and a half and double time respectively.

5.3.2.2 Operating Supervision

Operating supervision is the same as in previous sections.

5.3.2.3 Operating Consumables

Operating consumables costs were estimated as in previous sections.

5.3.2.4 Power

Power costs were estimated as in previous sections.

5.3.2.5 Maintenance

Maintenance costs were estimated as in previous sections.

5.3.2.6 Total Operating Costs

Operating costs are detailed in Table 11. A total operating cost of $3.10 per tonne milled is estimated.

TABLE 11 CASE A (500,000 tpa) - OPERATING COSTS

	Operating Labour			Operating Supervision			Operating Consumables			Power			Maintenance			Total		
	$	$/t	%	$	$/t	%	$	$/t	%	$	$/t	%	$	$/t	%	$	$/t	%
Crushing	26981	0.06	1.7	-	-	-	-	-	-	57097	0.11	3.7	190000	0.38	12.3	274078	0.55	17.7
Grinding	81690	0.16	5.3	-	-	-	329000	0.66	21.1	272734	0.55	17.6	100000	0.20	6.4	783424	1.57	50.4
Cassiterite Recovery	81690	0.16	5.3	-	-	-	-	-	-	58532	0.12	3.8	25000	0.05	1.6	165222	0.33	10.7
Concentrate Handling	81048	0.16	5.3	-	-	-	20000	0.04	1.3	5000	0.01	0.3	5000	0.01	0.3	111048	0.22	7.2
Administration + Overheads	81422	0.16	5.3	39000	0.08	2.5	60000	0.12	3.9	5000	0.01	0.3	31280	0.06	2.0	216702	0.32	14.0
Total	352831	0.70	22.9	39000	0.08	2.5	409000	0.82	26.3	398363	0.80	25.7	351280	0.70	22.6	1550474	3.10	100.0

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5.4 Case B - 1,100,000 Tonnes per Year

5.4.1 Capital Costs

Capital costs for crushing are the same as in previous cases. Capital costs downstream from the crusher are all assumed to be double those in Case A as downstream equipment is doubled in order to provide the required capacity. Table 12 shows estimates of capital costs.

Table 12 - 1,100,000 tpa Capital Costs

Section	$
Crushing	1,810,000
Grinding	2,142,000
Cassiterite Recovery	1,356,000
Concentrate Handling and Miscellaneous	4,110,000
Contingency (20% of total of above items)	1,883,000
Total	11,300,000

5.4.2 Operating Costs

5.4.2.1 Operating Labour

5.4.2.1.1 Manning

Operator numbers for the crusher are increased by seven due to extending operating hours. Two operators per shift have been included as no assistance from day crew personnel is available on afternoon and night shifts.

Operating strength in the grinding and cassiterite recovery sections can be maintained at previous levels.

Day crew personnel should be increased in number to a total of eight. This is necessary to cope with increased concentrate production and increases in clean-up and housekeeping duties.

5.4.2.1.2 Costs

Costs are estimated as in previous sections.

5.4.2.2 Operating Supervision

Non-technical supervision is justified to assist in the operation due to the increases in manning levels. A general foreman with a salary of $20,000 (plus 30 per cent as overheads) has been included.

5.4.2.3 Operating Consumables

Operating consumables costs have been estimated as previously described.

5.4.2.4 Power

Power costs were estimated as described in previous sections.

5.4.2.5 Maintenance

Maintenance costs were estimated as in previous sections.

5.4.2.6 Total Operating Costs

Operating costs are detailed in Table 13. A total operating cost of $2.86 per tonne treated is estimated.

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TABLE 13 CASE B (1,100,000 tpa) - OPERATING COSTS

	Operating Labour			Operating Supervision			Operating Consumables			Power			Maintenance			Total		
	$	$/t	%	$	$/t	%	$	$/t	%	$	$/t	%	$	$/t	%	$	$/t	%
Crushing	163380	0.15	5.2	-	-	-	-	-	-	128469	0.12	4.1	418000	0.38	13.3	709849	0.65	22.6
Grinding	81690	0.07	2.6	-	-	-	726000	0.66	23.1	545468	0.50	17.3	220000	0.20	7.0	1573158	1.43	50.0
Cassiterite Recovery	81690	0.07	2.6	-	-	-	-	-	-	117064	0.11	3.7	55000	0.05	1.7	253754	0.23	8.0
Concentrate Handling	129677	0.12	4.2	-	-	-	44000	0.04	1.4	11000	0.01	0.3	11000	0.01	0.3	195677	0.18	6.2
Administration + Overheads	136931	0.12	4.4	65000	0.06	2.1	132000	0.12	4.2	11000	0.01	0.3	68816	0.06	2.2	413747	0.37	13.2
Total	593368	0.53	19.0	65000	0.06	2.1	902000	0.82	28.7	813001	0.75	25.7	772816	0.70	24.5	3146185	2.86	100.0

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6. CONCLUSIONS

Blue Tier mineralisation, although low grade, contains coarse, easily liberated cassiterite which can be efficiently recovered by simple, low cost gravity processes to a concentrate of marketable grade.

The ease and simplicity of processing ensures that the mineralisation can be treated with low levels of capital expenditure and at very low treatment costs.

Table 14 summarises the three cases examined and shows:

1. As annual throughput increases, manning levels have to be increased.

2. As annual throughput increases, capital expenditure must be increased.

3. As annual throughput increases, the total annual operating cost is increased but the unit cost per tonne treated is reduced.

TABLE 14 - SUMMARY OF OPERATIONS

		BASE CASE	CASE A	CASE B
Annual Treatment Rate (t)		350,000	500,000	1,100,000
Mine Life (yr)		10	7	3
Tin production (t/yr)		841	1202	2643
Manning	- Award	11	14	24
	- Staff	1	1	2
	- Total	12	15	26
Capital Cost ($)		6,343,000	6,737,000	11,300,000
Operating Cost ($/yr)		1,099,939	1,550,474	3,146,185
	($/t)	3.16	3.10	2.86

350,000
@
0.27%
= 945
Prodn
841
∴ Recovery
89%

29% 28%

7. RECOMMENDATIONS

1. Only a 'skin and bones' operation can be profitable at Blue Tier. This should be a conscious thought in any future planning or investigations.

2. As exploitation of the deposit is dependant upon an ability to dispose of tailings from the concentrator a suitable site and method for tailing disposal should be investigated.

3. Any mineralisation encountered in the area in the future should be subjected to metallurgical testwork to assess its performance in the proposed concentrator.

4. Any future definitive investigations of the deposit should include a study of the relationship between revenue and base metal impurity levels in concentrates under the likely terms of a contract for Blue Tier concentrates.

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8. ACKNOWLEDGEMENTS

Particular thanks are due to G. Bryan and E. Prince for their efforts and contributions. They conducted the bulk of the Renison testwork and their knowledge and experience proved invaluable during the project.

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APPENDIX I

HEAVY MEDIUM SEPARATION

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FIGURE AI-1

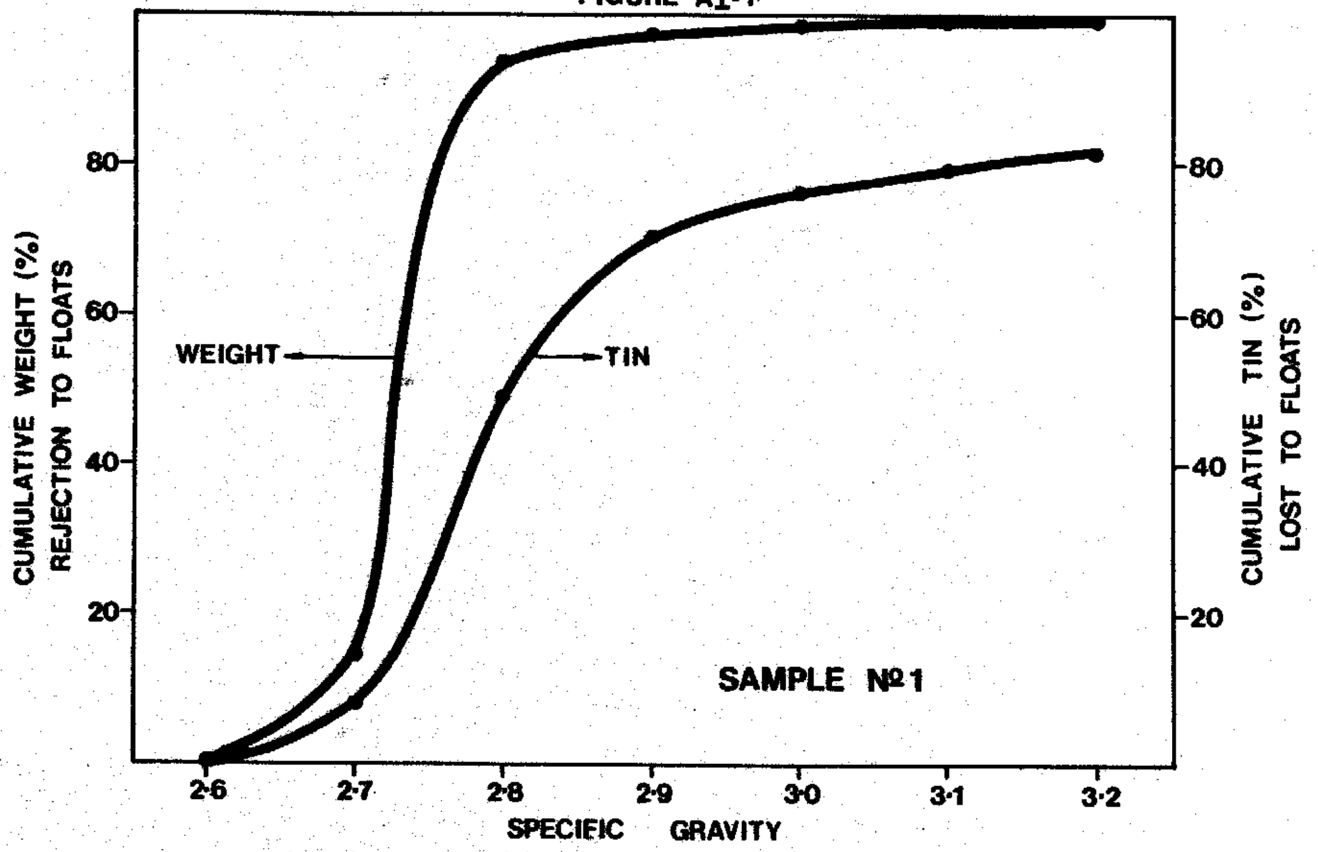
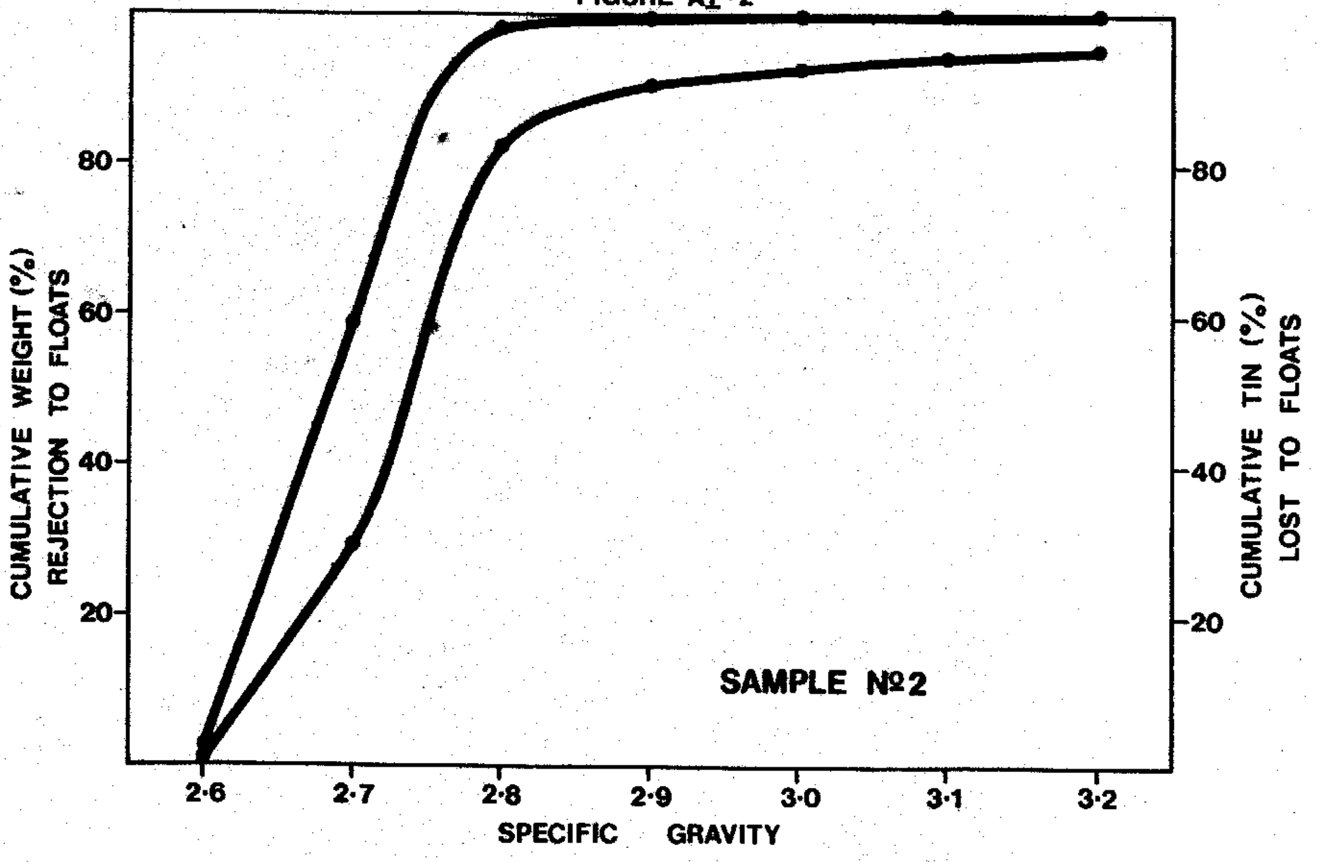


FIGURE AI-2



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AI-1 Testwork

Approximately 30 kilograms of -12.7 + 0.5mm material from each of bulk samples No. 1 and 2 were submitted to AMDEL for Heavy Medium Separation Feasibility Study testwork.

AI-2 Results

Results from the AMDEL are shown in tables AI-1 and AI-2.

Table AI-1. H.M.S. Feasibility Study Results for Sample No. 1

Sp.gr. Product	wt%	Assay % Sn	Distribution % Sn
<2.6	0.28	0.027	0.04
2.6 - 2.7	13.76	0.100	7.35
2.7 - 2.8	79.18	0.098	41.43
2.8 - 2.9	4.18	0.98	21.87
2.9 - 3.0	1.45	0.73	5.65
3.0 - 3.1	0.42	1.34	3.01
3.1 - 3.2	0.22	2.03	2.38
3.2 - 3.3	0.14	2.40	1.79
>3.3	0.37	8.34	16.48

FIGURE AI-3

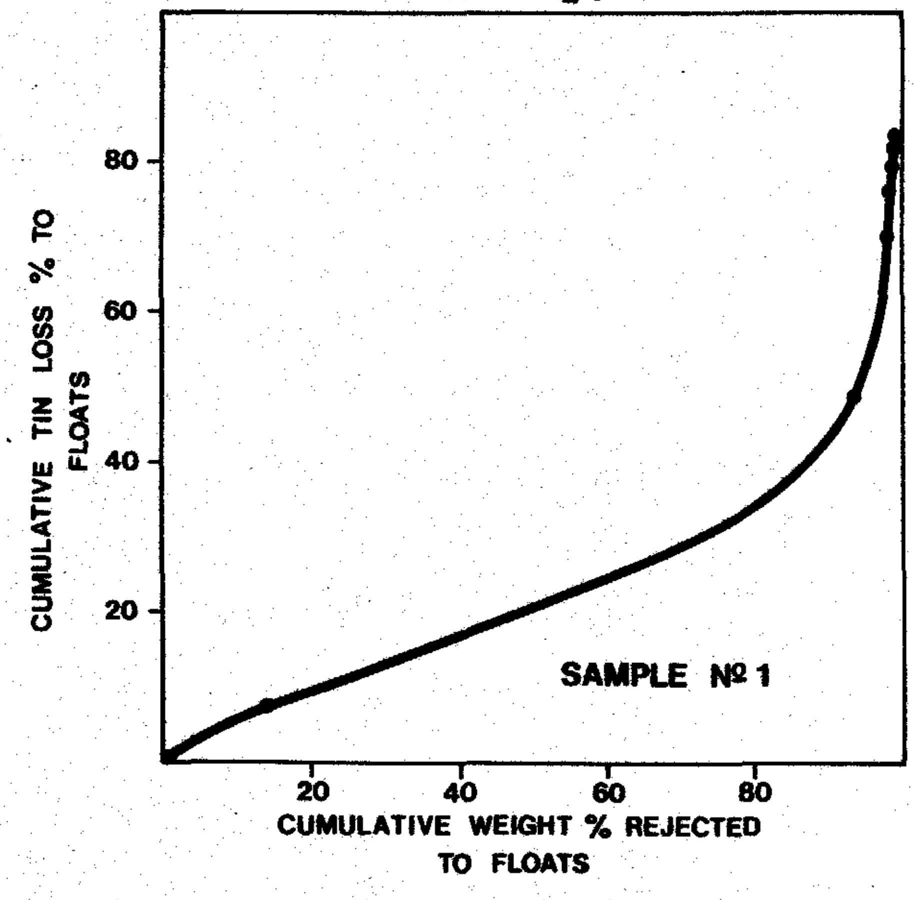

FIGURE AI-4

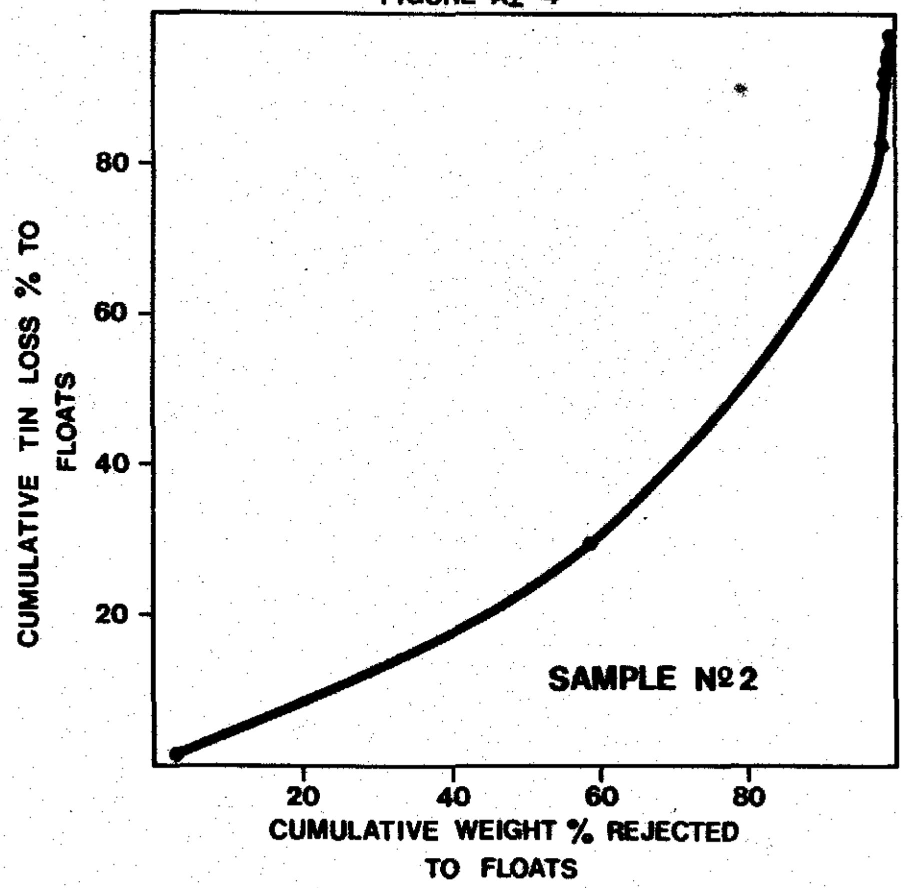

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Table AI-2 H.M.S. Feasibility Study Results for Sample No. 2

Sp.gr. Product	Wt%	Assay % Sn	Distribution % Sn
<2.6	2.63	0.11	1.06
2.6 - 2.7	55.76	0.14	28.73
2.7 - 2.8	39.58	0.36	52.45
2.8 - 2.9	1.68	1.34	8.29
2.9 - 3.0	0.14	4.67	2.41
3.0 - 3.1	0.08	4.47	1.25
3.1 - 3.2	0.05	7.37	1.46
3.2 - 3.3	0.02	8.41	0.56
>3.3	0.06	16.1	3.79

AI-3 Discussion of Results

Figures AI-1 and AI-2 show plots of weight per cent and tin per cent rejected to floats versus specific gravity for the two samples tested. At all specific gravities an unacceptable loss of tin occurs for the levels of weight rejection achieved.

This is further emphasised by Figures AI-3 and AI-4 where weight rejection is plotted versus tin loss. These results show H.M.S. to be undesirable as a preconcentration stage unless high losses of recoverable mineral to floats are tolerated.

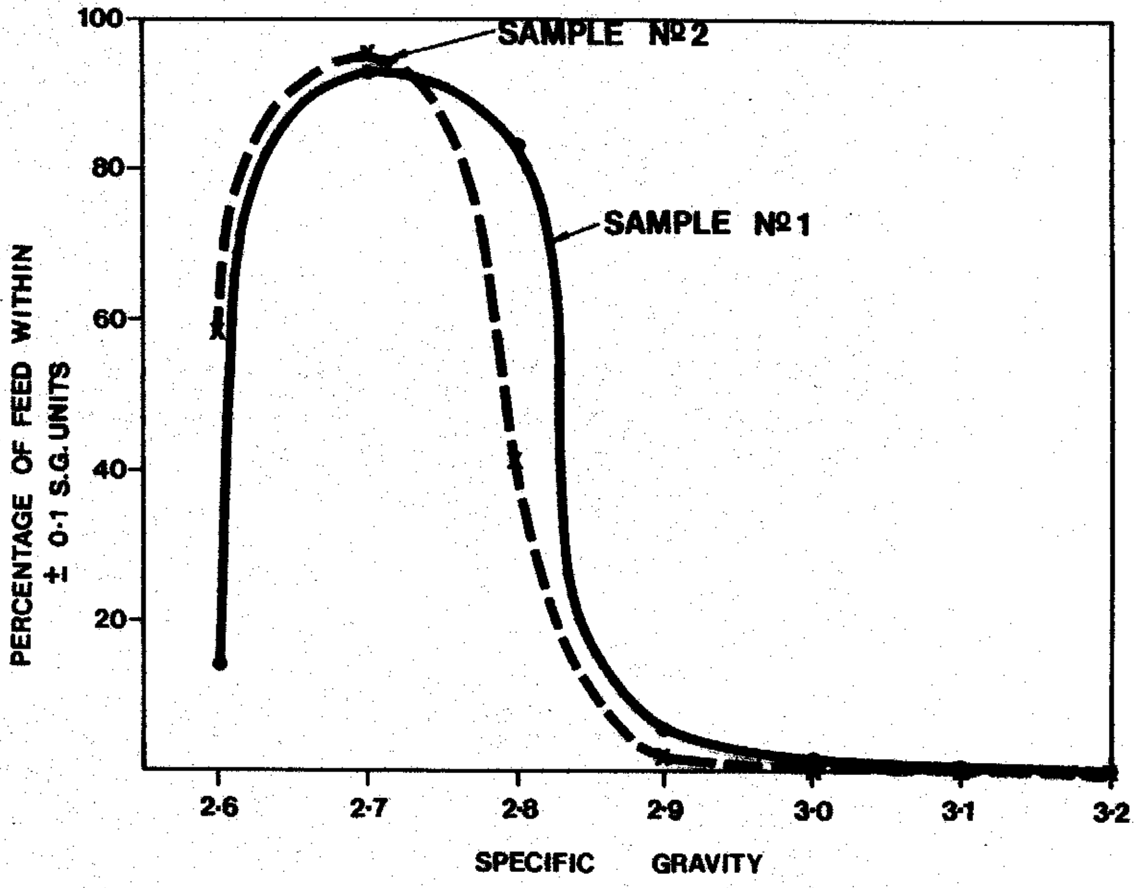
SAMPLE №2
SAMPLE №1
100
80
60
40
20
PERCENTAGE OF FEED WITHIN ± 0·1 S.G. UNITS
2·6
2·7
2·8
2·9
3·0
3·1
3·2
SPECIFIC GRAVITY

FIGURE AI-5

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Figure AI-5 shows a plot of the proportion of the feed within plus or minus 0.1 specific gravity units versus specific gravity for the two samples. These results demonstrate that even if high tin losses were acceptable thereby making H.M.S. feasible, extremely close control of operating density would be required to achieve satisfactory results. Such control would be difficult and it is dubious whether it could be practically maintained.

AI-4 Conclusion

For the reasons outlined above, heavy media separation was not considered as a feasible preconcentration operation.

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APPENDIX II

HEAVY LIQUID TEST RESULTS

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AII-1 **Heavy Liquid Test Results**

Results from heavy liquid testwork carried out on the 4 bulk samples and drill core samples are shown schematically in Figures AII-1 to AII-15. An examination of the figures shows there is considerable variation in tin distribution between size fractions, between specific gravity products for the total sample and between specific gravity products within individual size fractions of the sample.

AII-2 **Liberation Size**

As size is reduced liberation improves. At a certain size almost maximum liberation is achieved and as further reductions are made below this size, liberation is improved to a much lesser extent. Determination of this critical size enables a liberation size and hence a grind size to be established.

AII-3 **Required Size of Grind**

Data from Figures AII-1 to AII-15 was plotted as shown in Figures AII-16 to AII-30. In these figures mean particle size is plotted against tin distribution to the product with a specific gravity greater than 3.3 within that size fraction. Liberation sizes ranged from 190µm to 560µm. Results for the fifteen samples are shown in Table AII-1.

BULK SAMPLE Nº 1

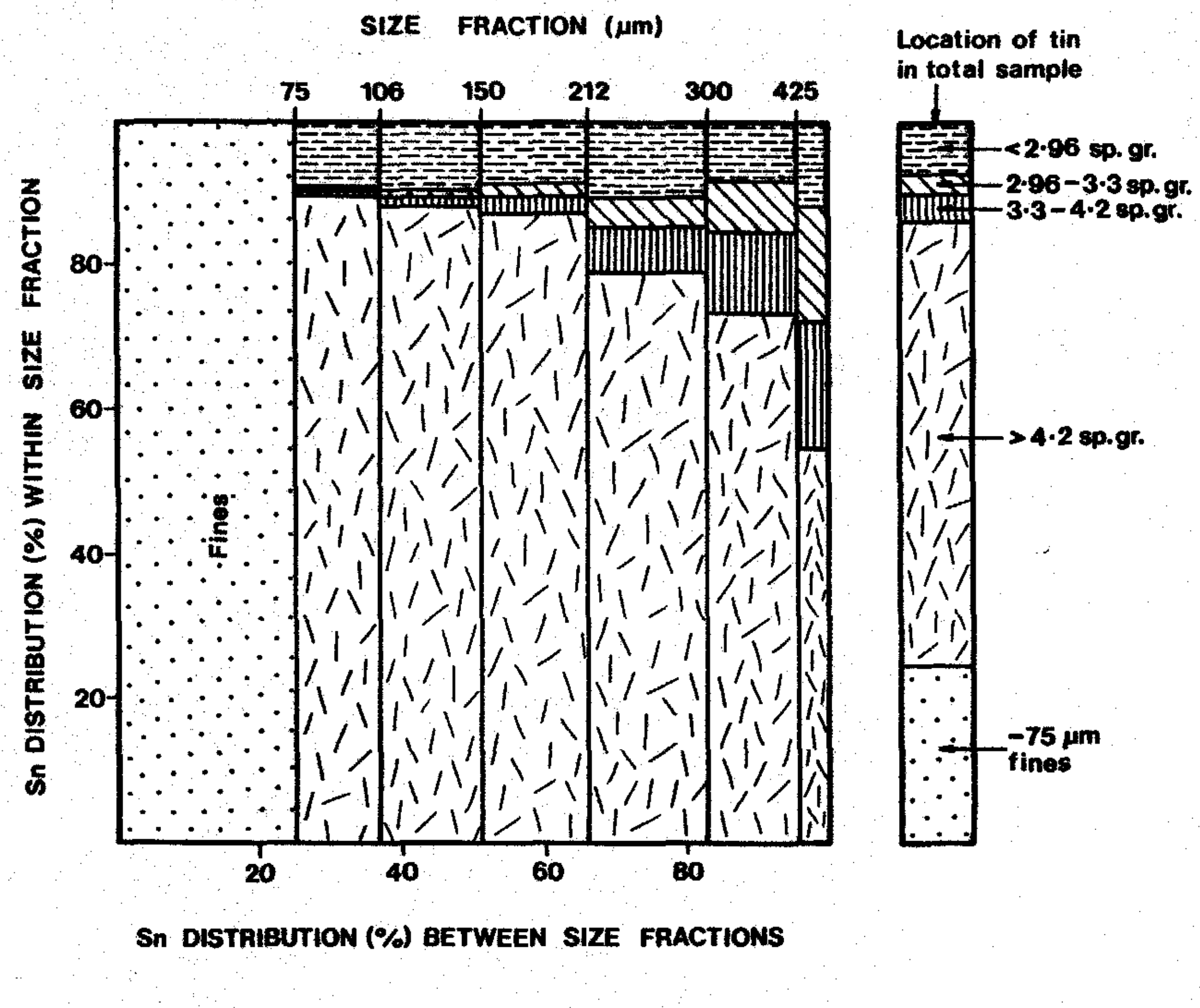

FIGURE AII-1

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BULK SAMPLE №2

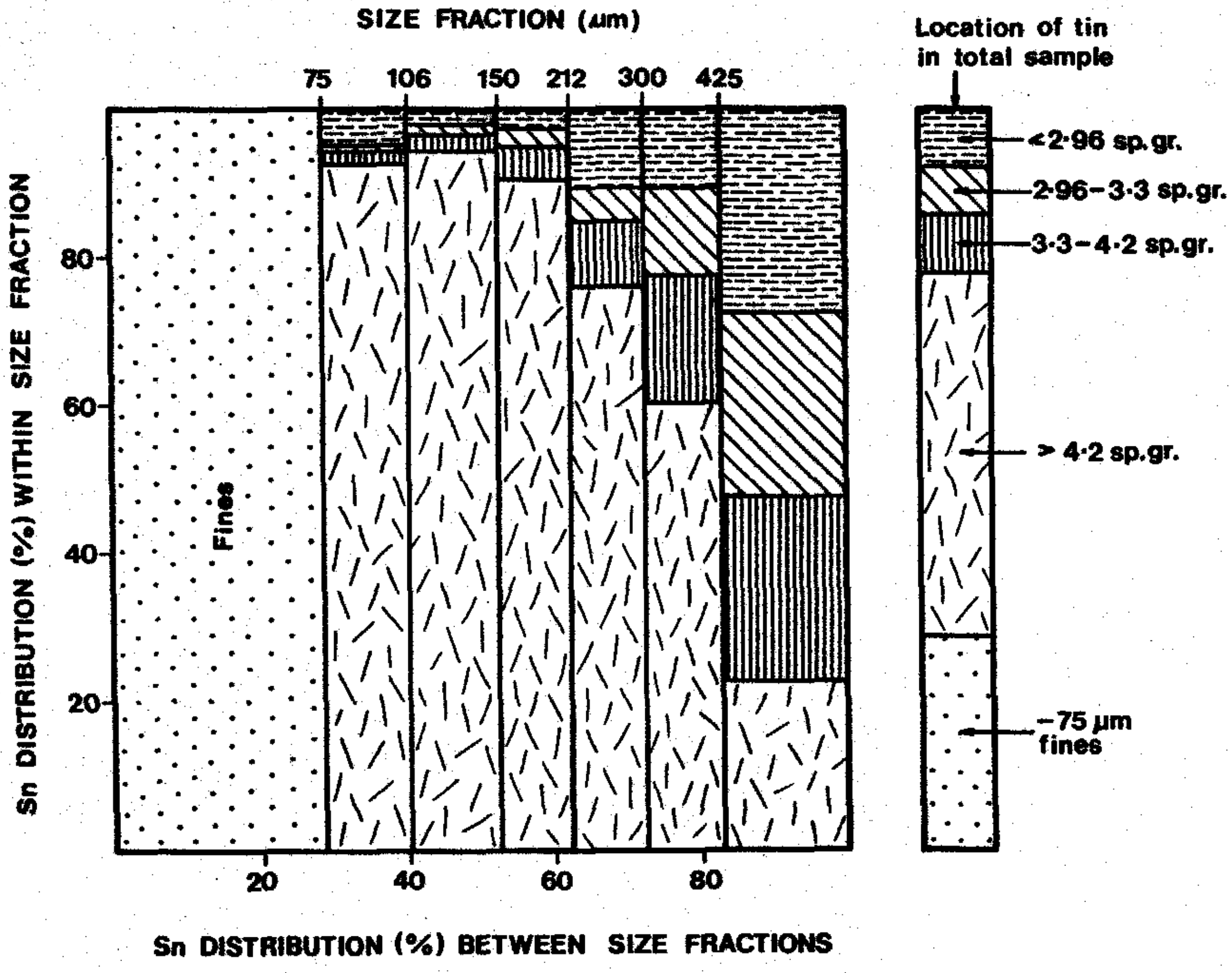

FIGURE AII-2

066

BULK SAMPLE Nº 3A

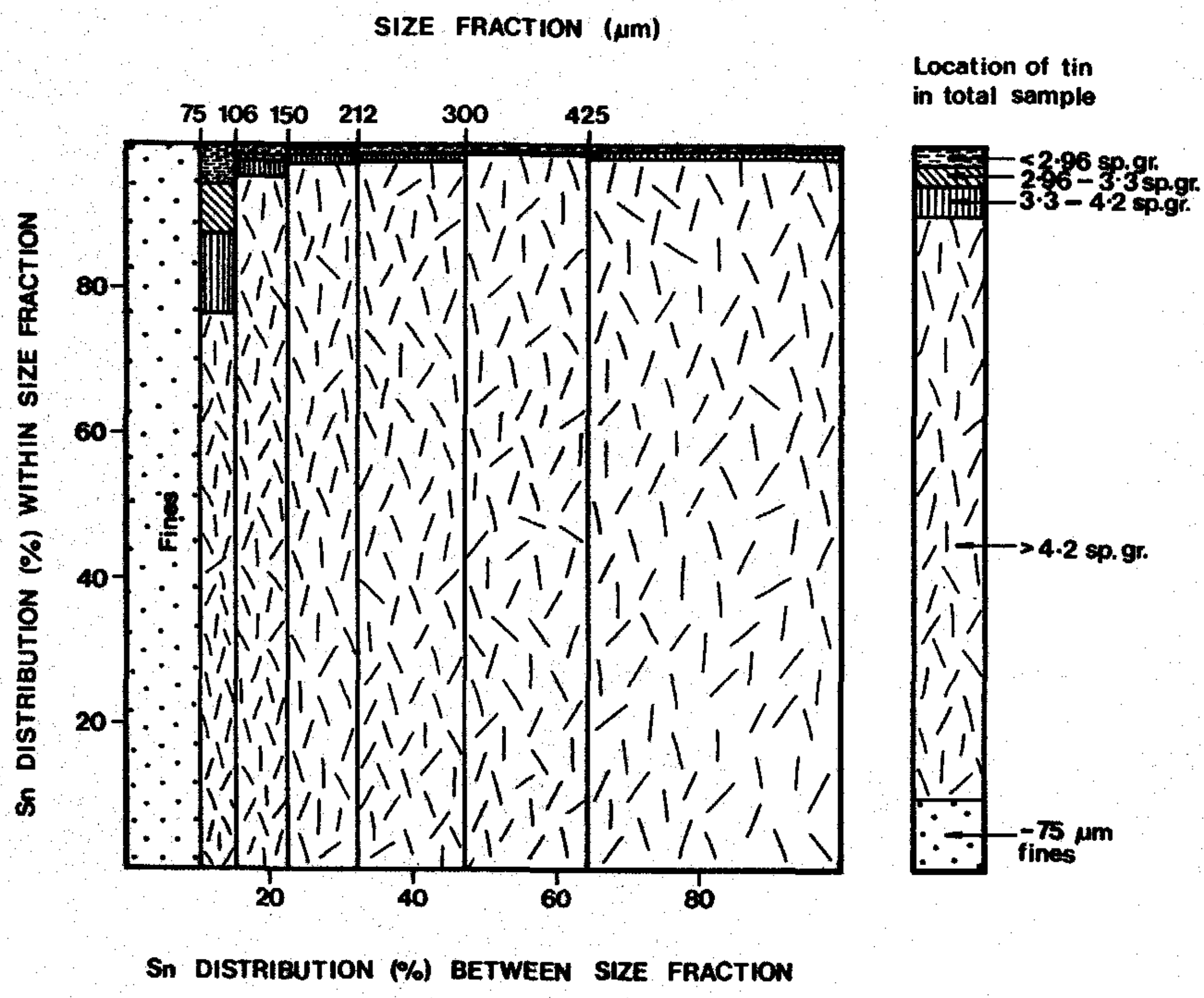

FIGURE AII-3

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067

BULK SAMPLE Nº 3B

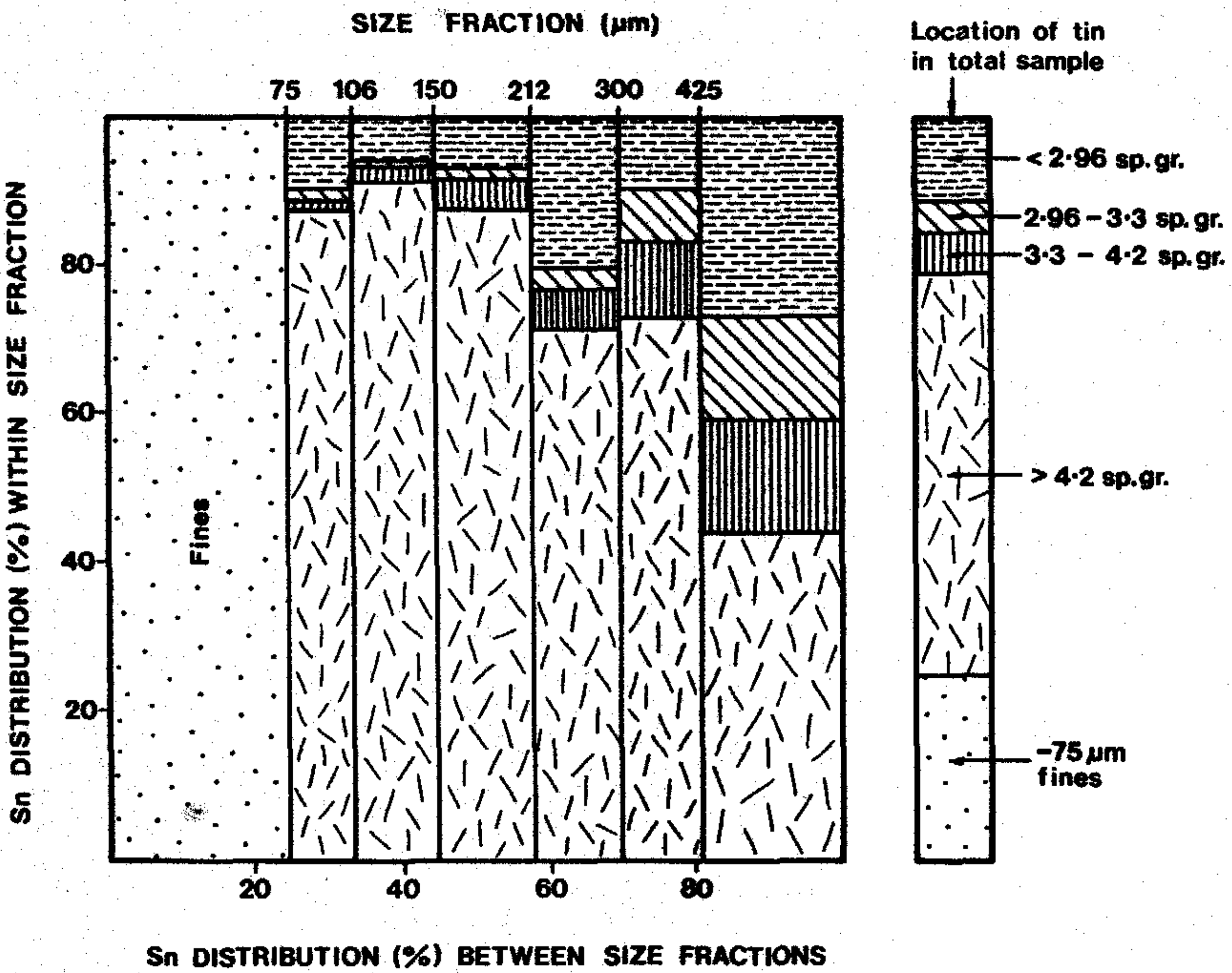

FIGURE AII-4

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067

BT 48 (49·5 – 56·5)

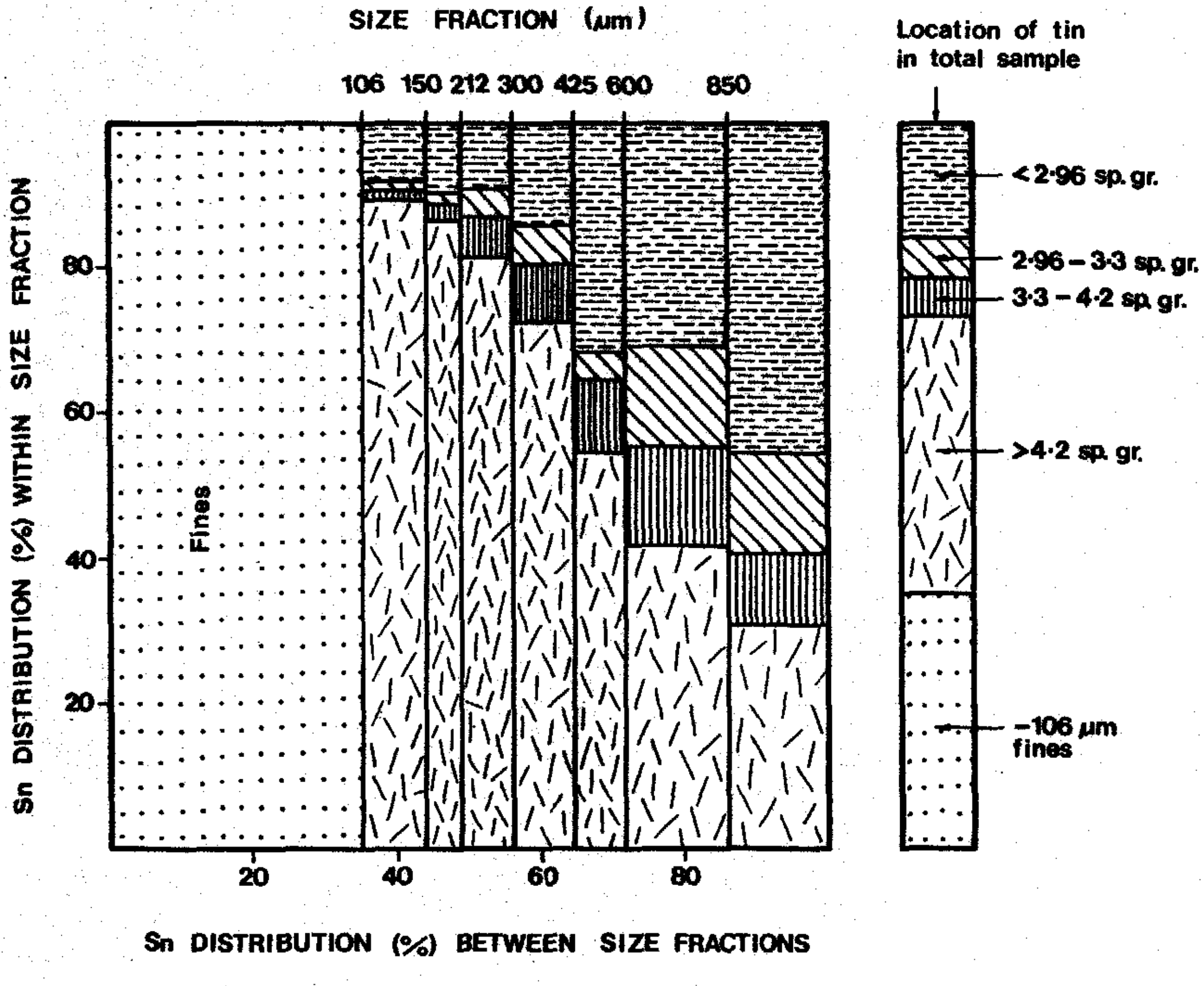

FIGURE AII-5

068

903071

BT 59A (0 – 13)

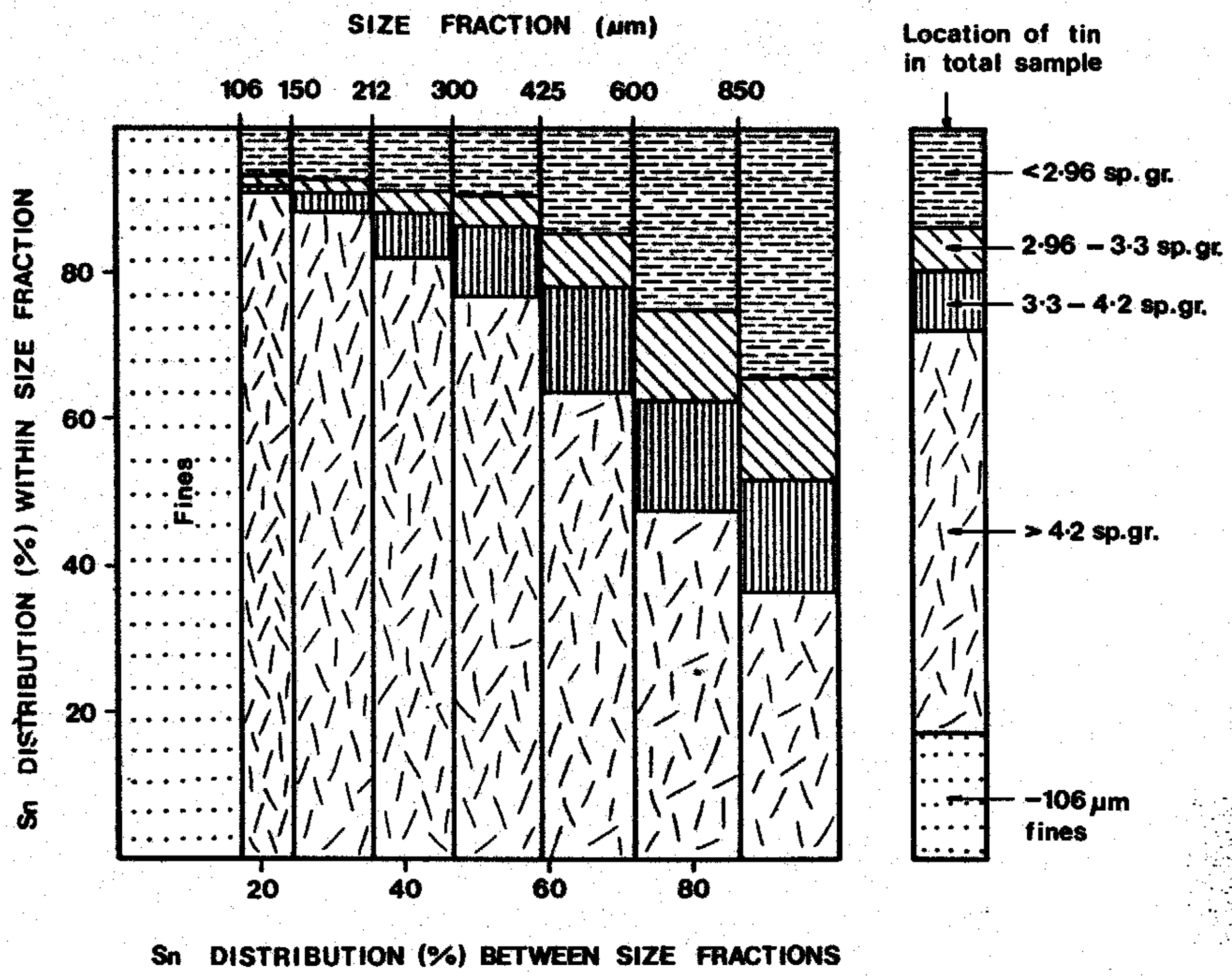

FIGURE AII-6

069

BT 59B (25-34)

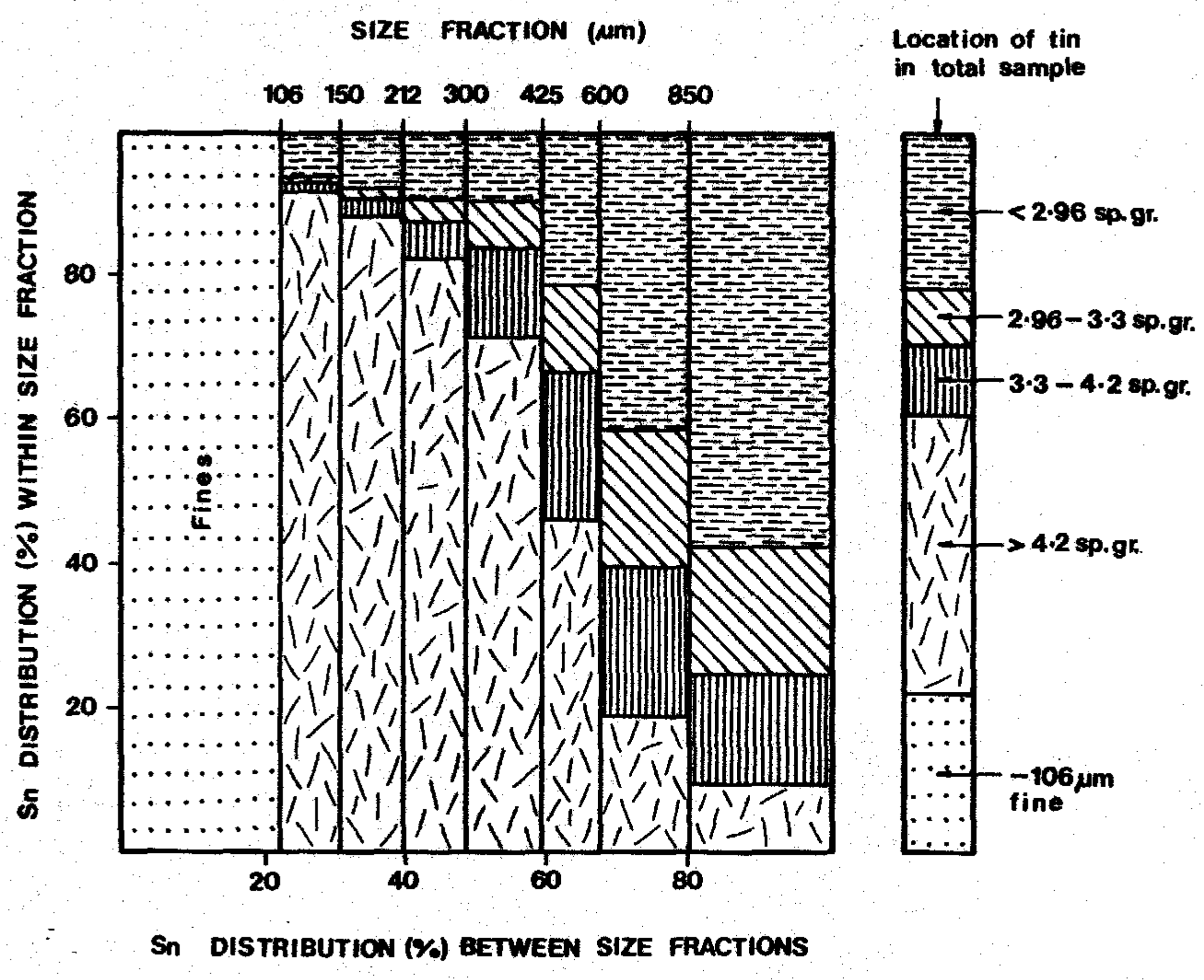

FIGURE AII-7

070

BT 69 (42·6 - 53·6)

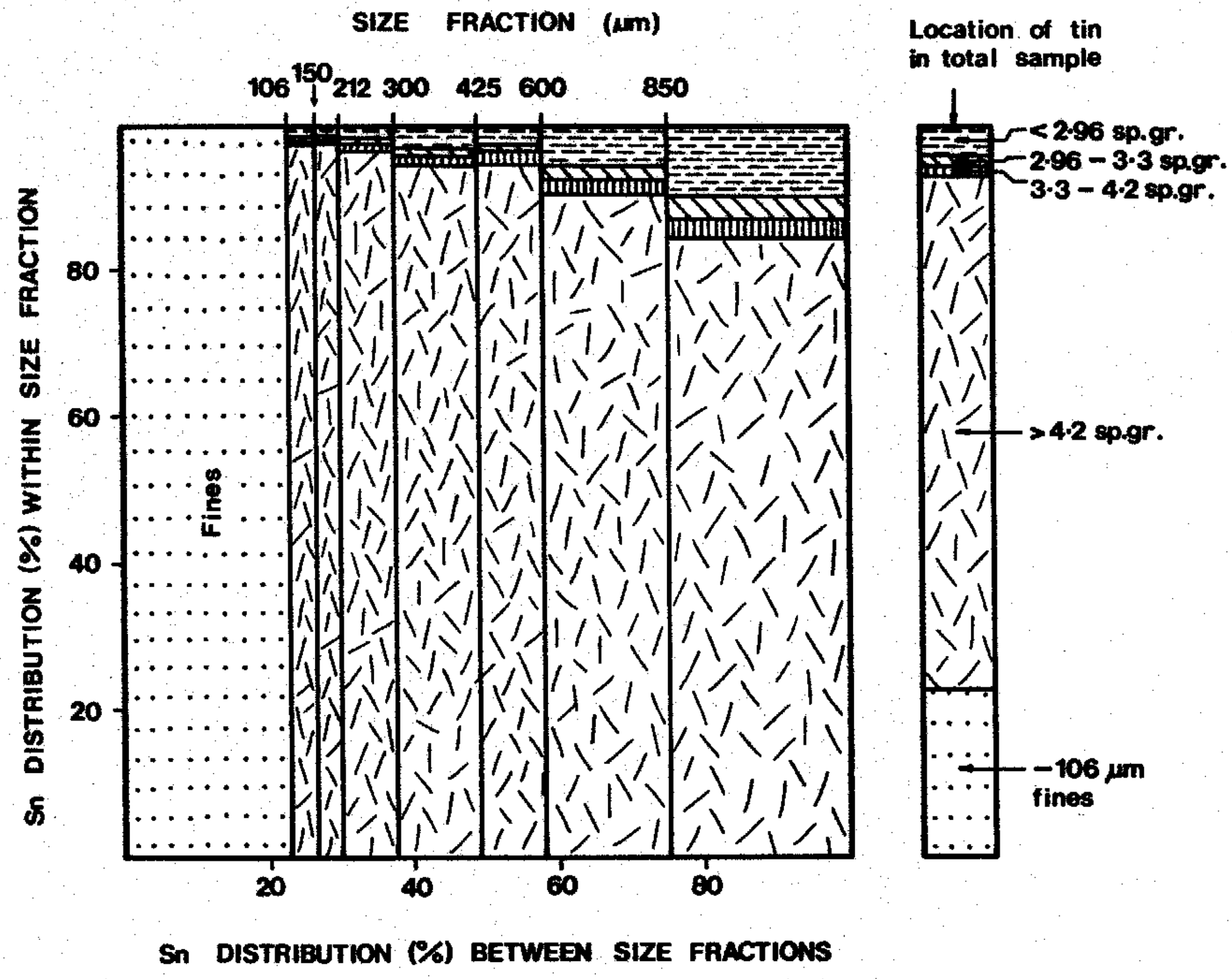

FIGURE AII-8

071

BT 71 (0-71)

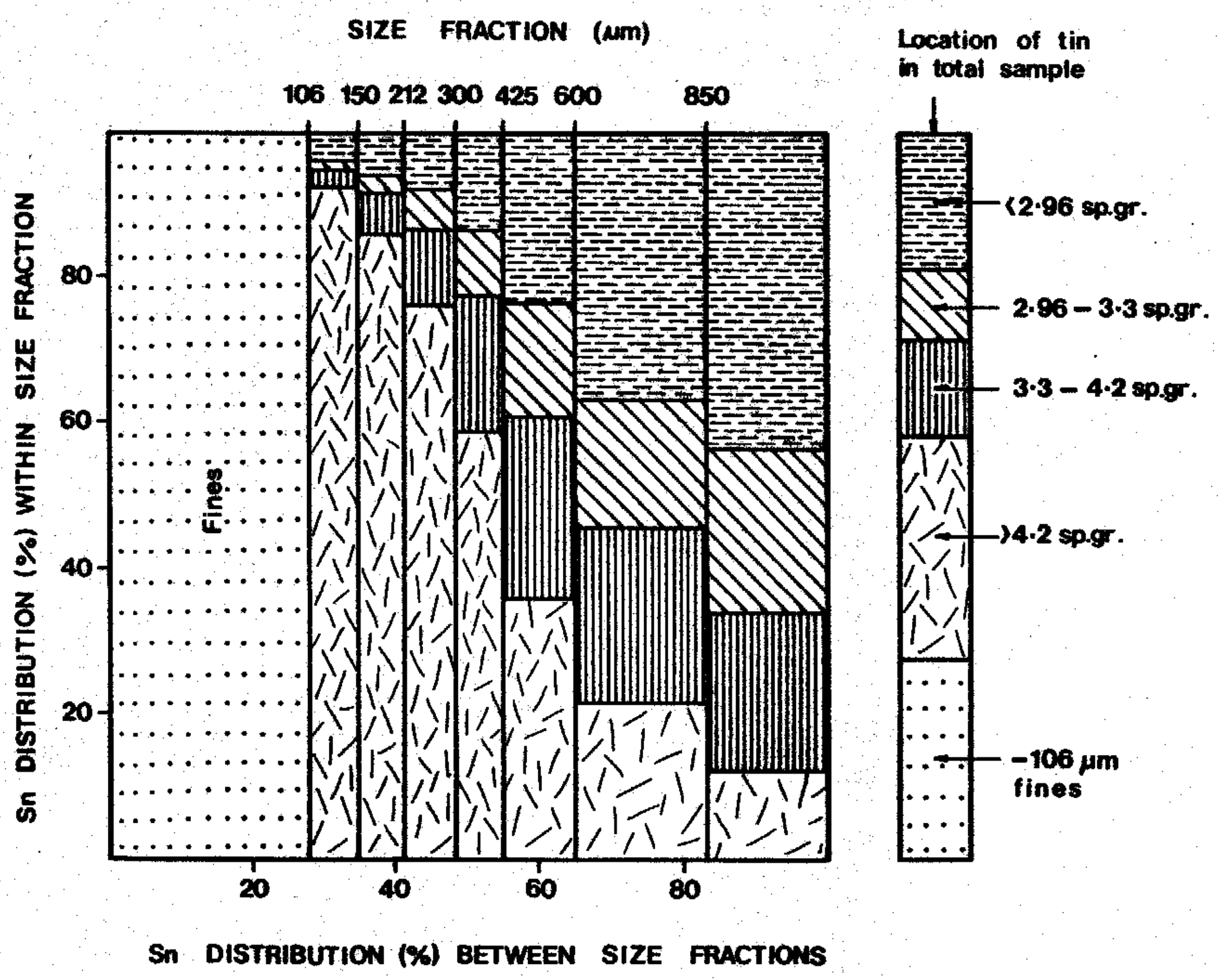

FIGURE AII-9

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072

BT 81A (127·3 – 135·9)

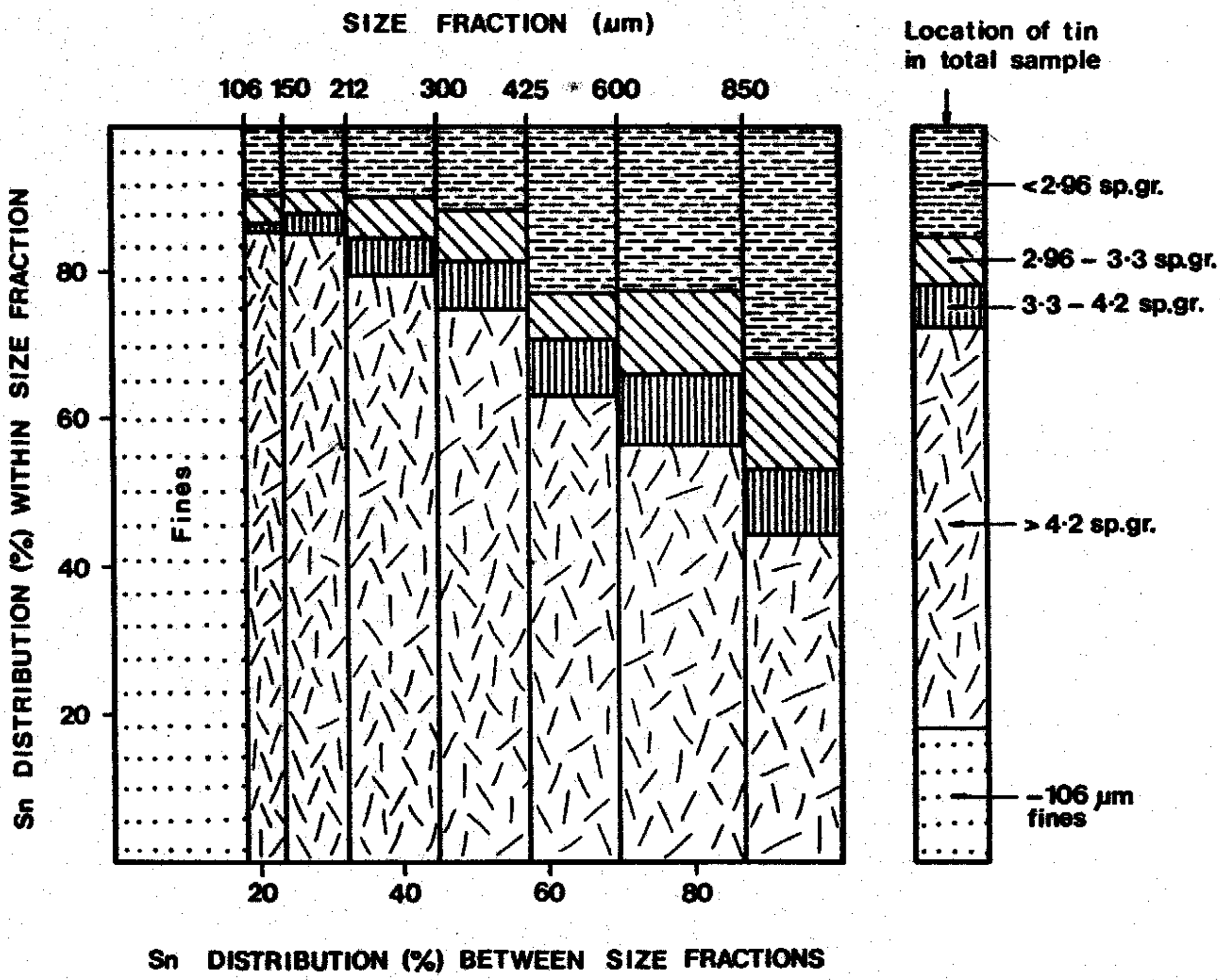

FIGURE AII-10

BT 81B (143·4 – 148·4)

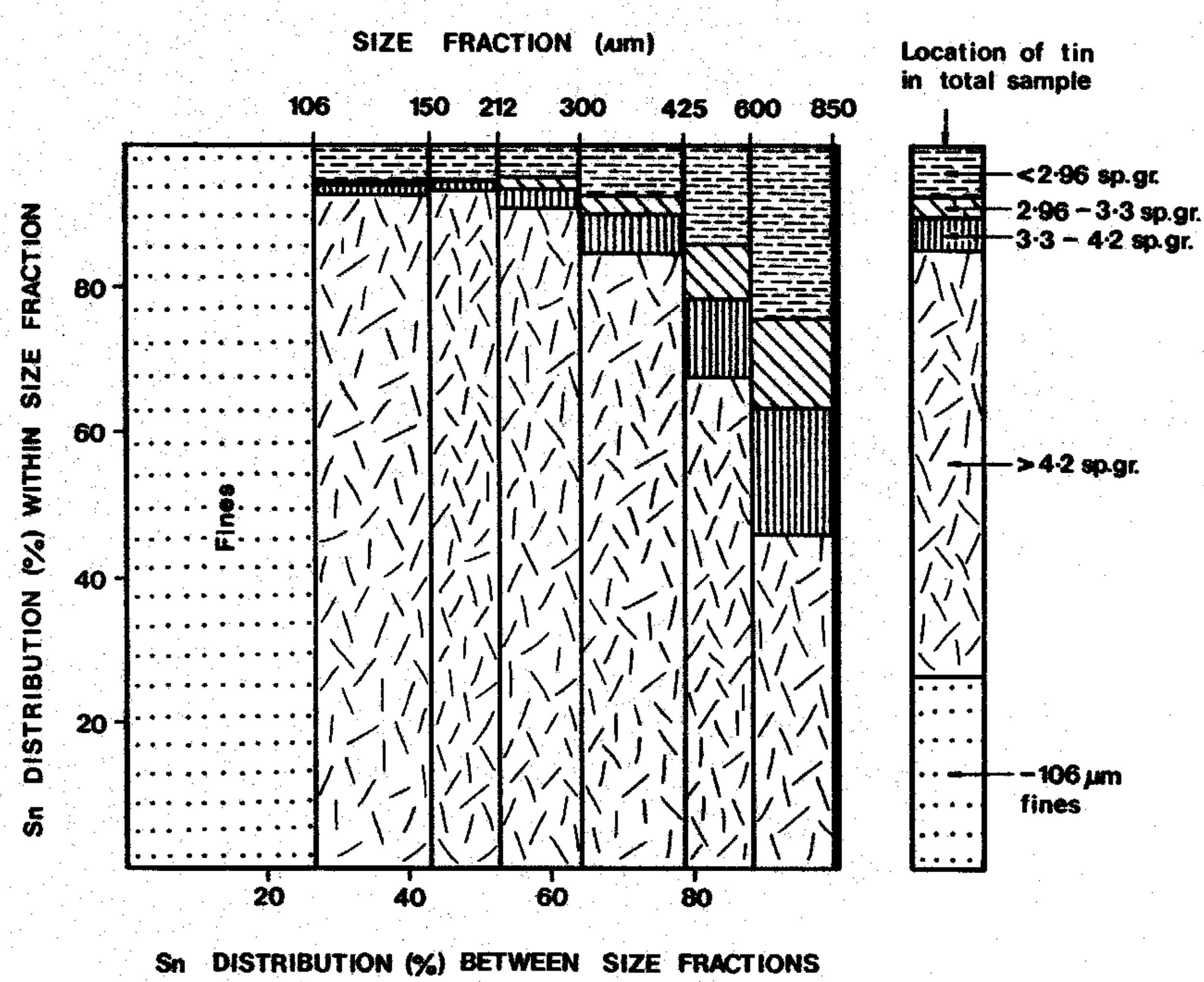

FIGURE AII·11

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074

BT 82 (60-67)

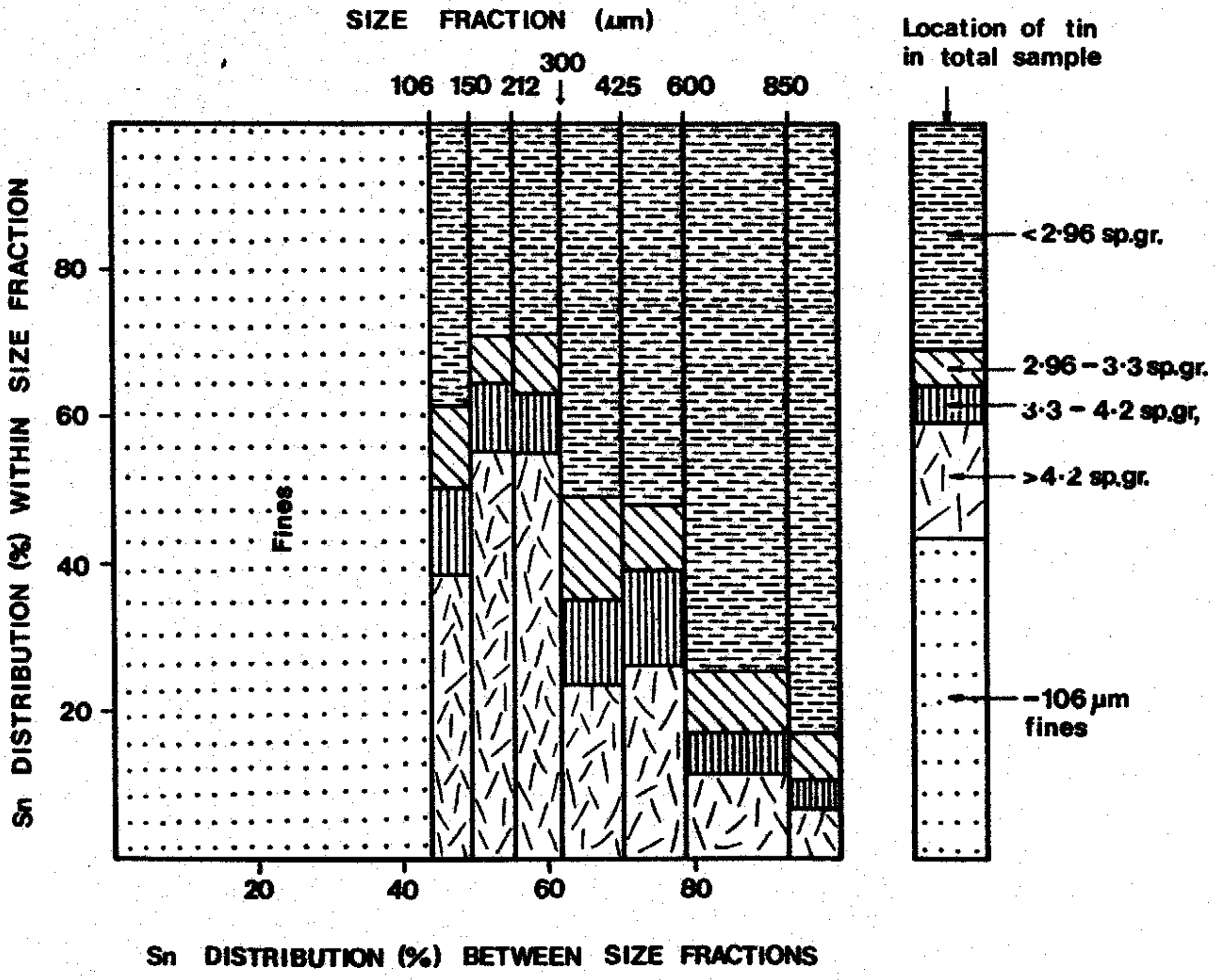

FIGURE AII-12

075

BT 84 (5-27)

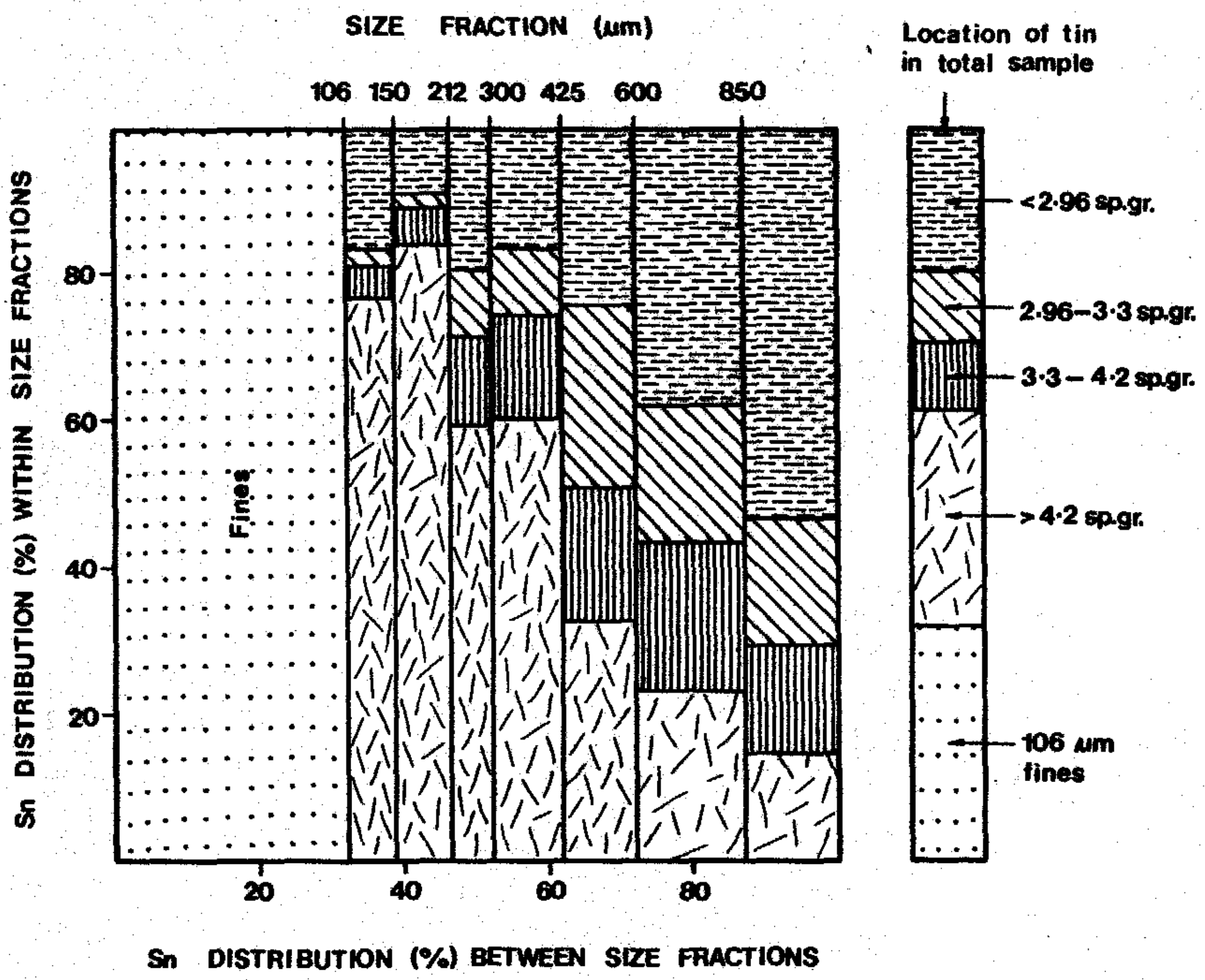

FIGURE AII-13

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076

BT 86 (57–65·2)

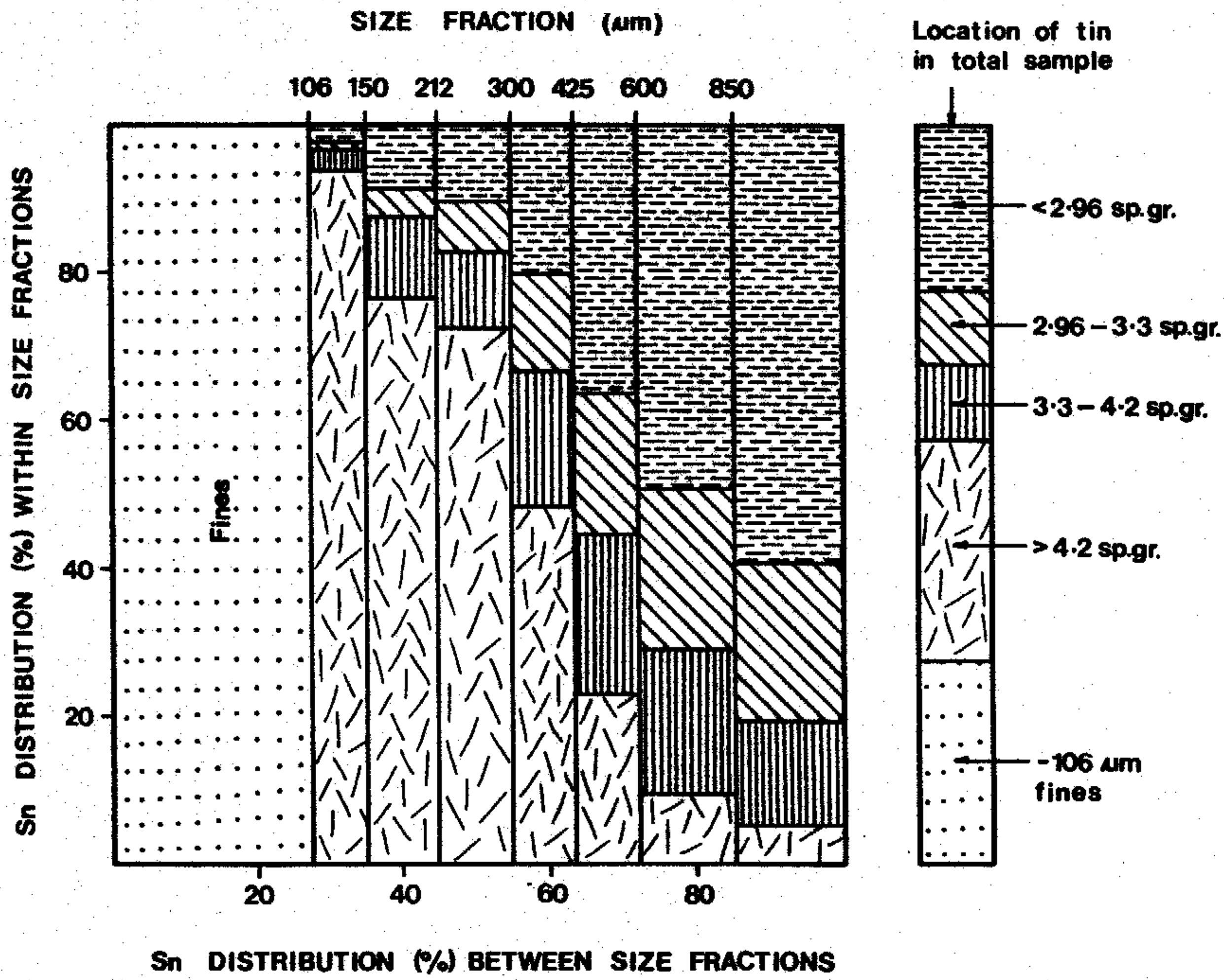

FIGURE AII-14

BT 89 (0–25·6)

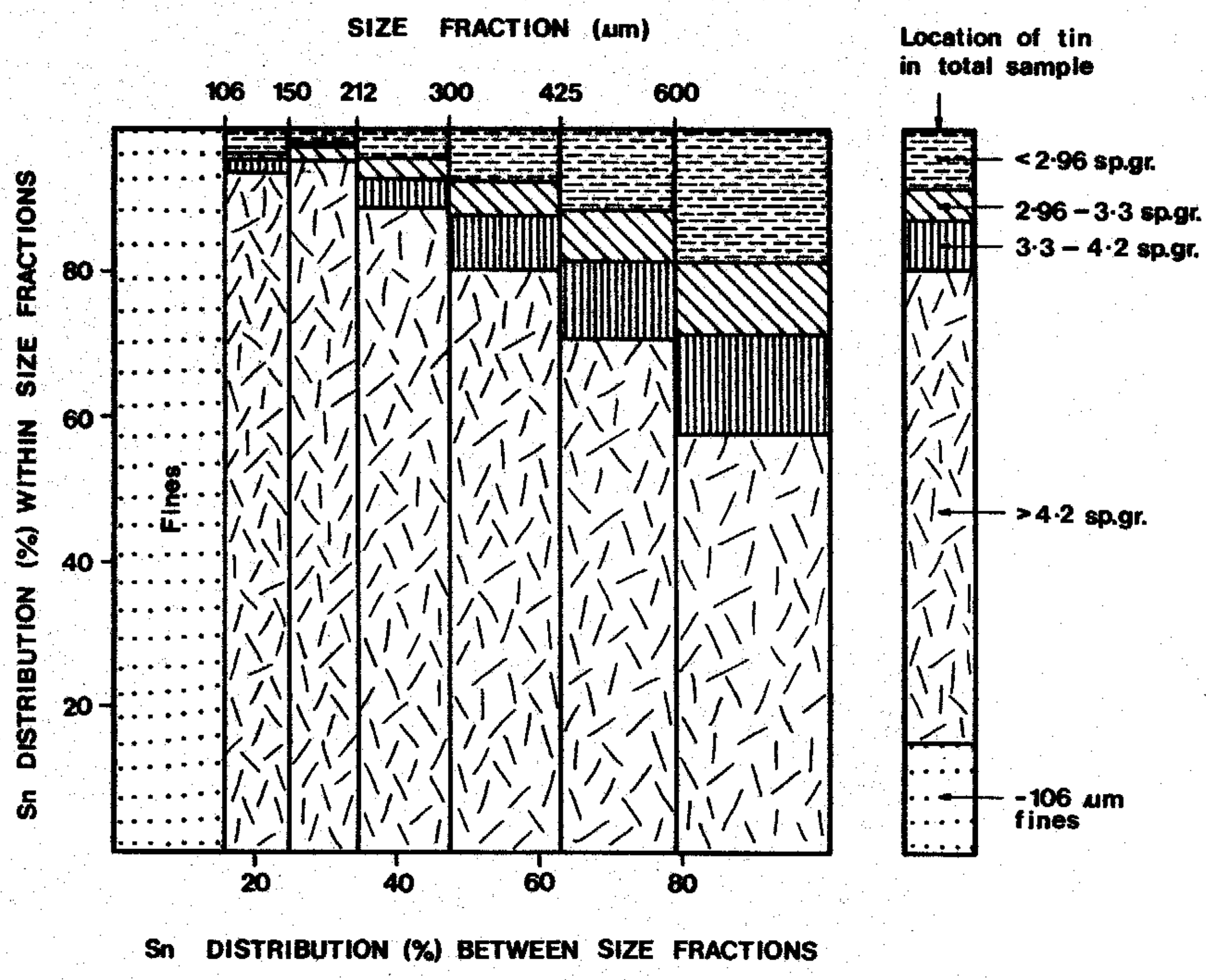

FIGURE AII-15

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078

BULK SAMPLE Nº 1

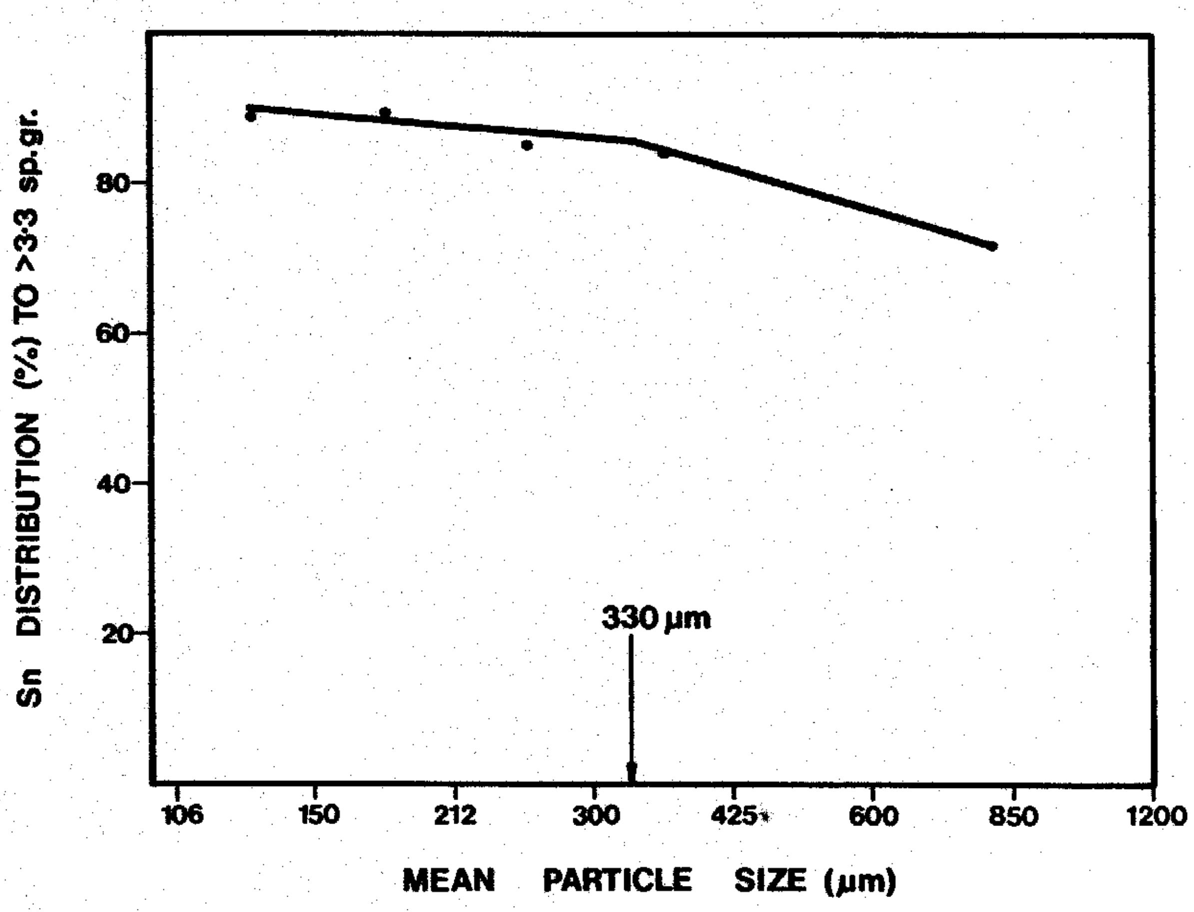

FIGURE AII-16

079

BULK SAMPLE Nº 2

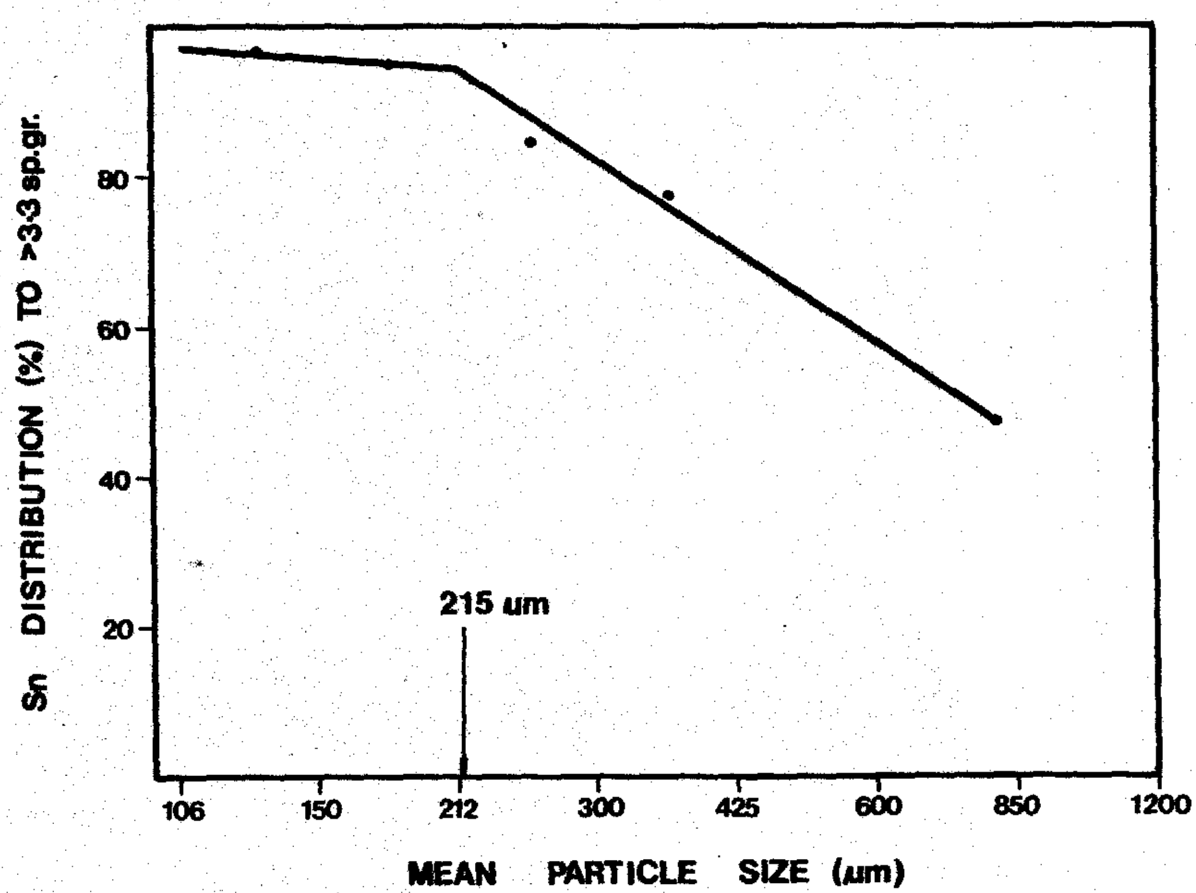

FIGURE AII-17

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080

BULK SAMPLE Nº 3A

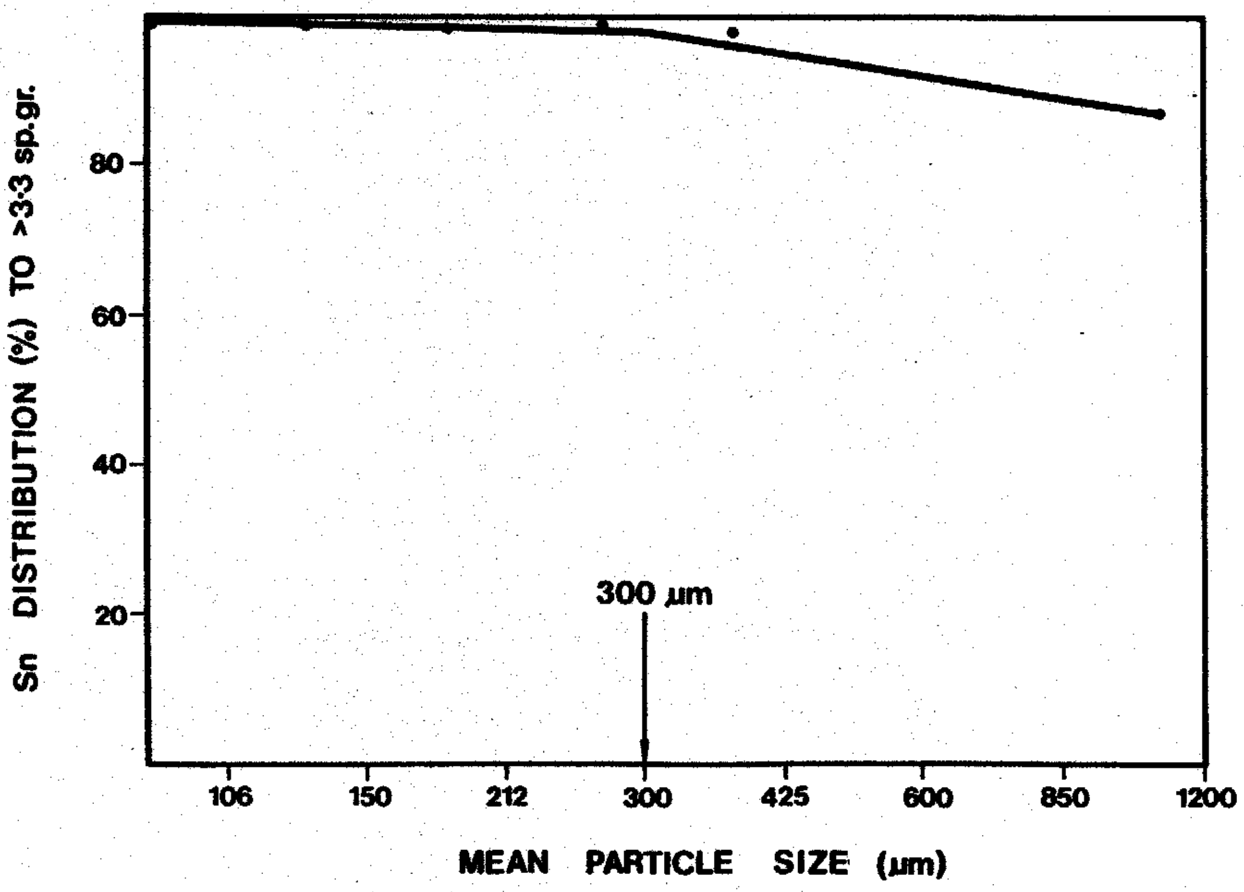

FIGURE A II-18

BULK SAMPLE Nº 3B

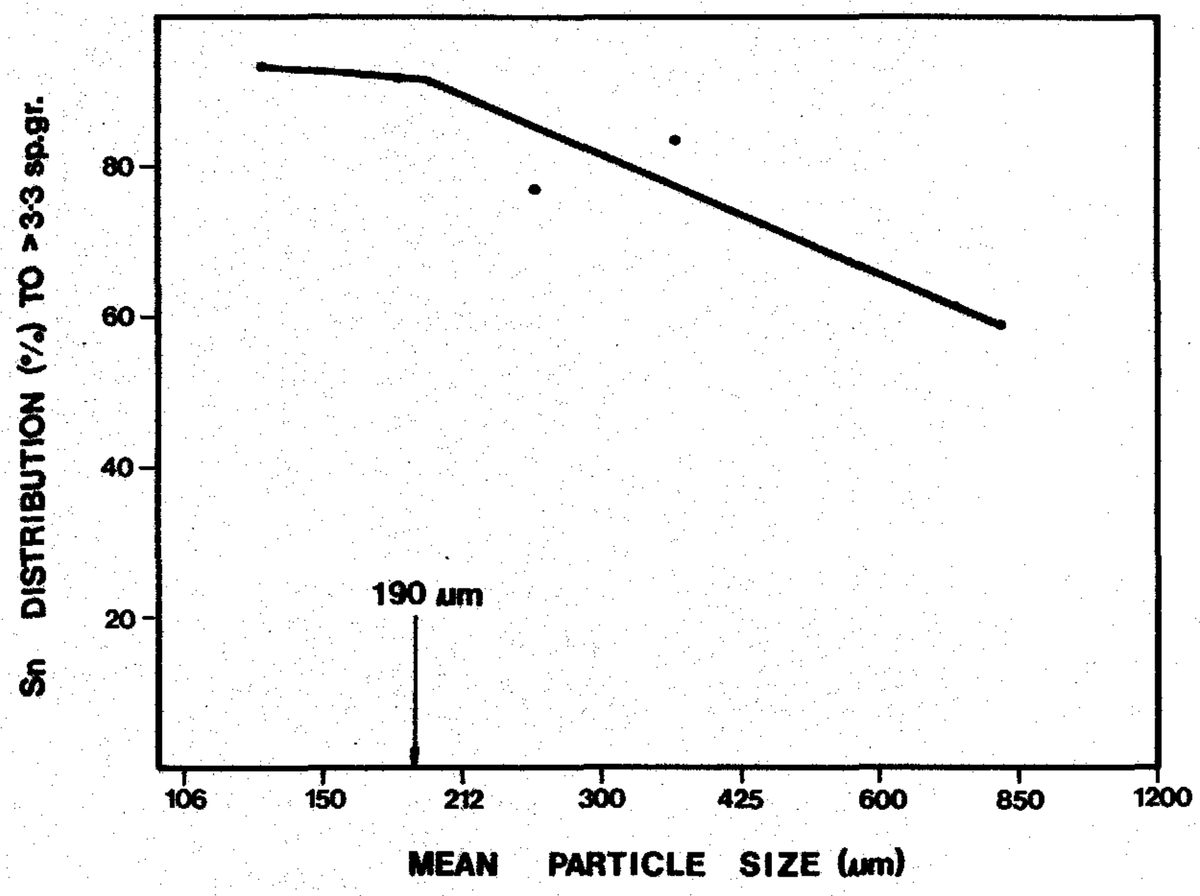

FIGURE AII-19

082

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BT 48

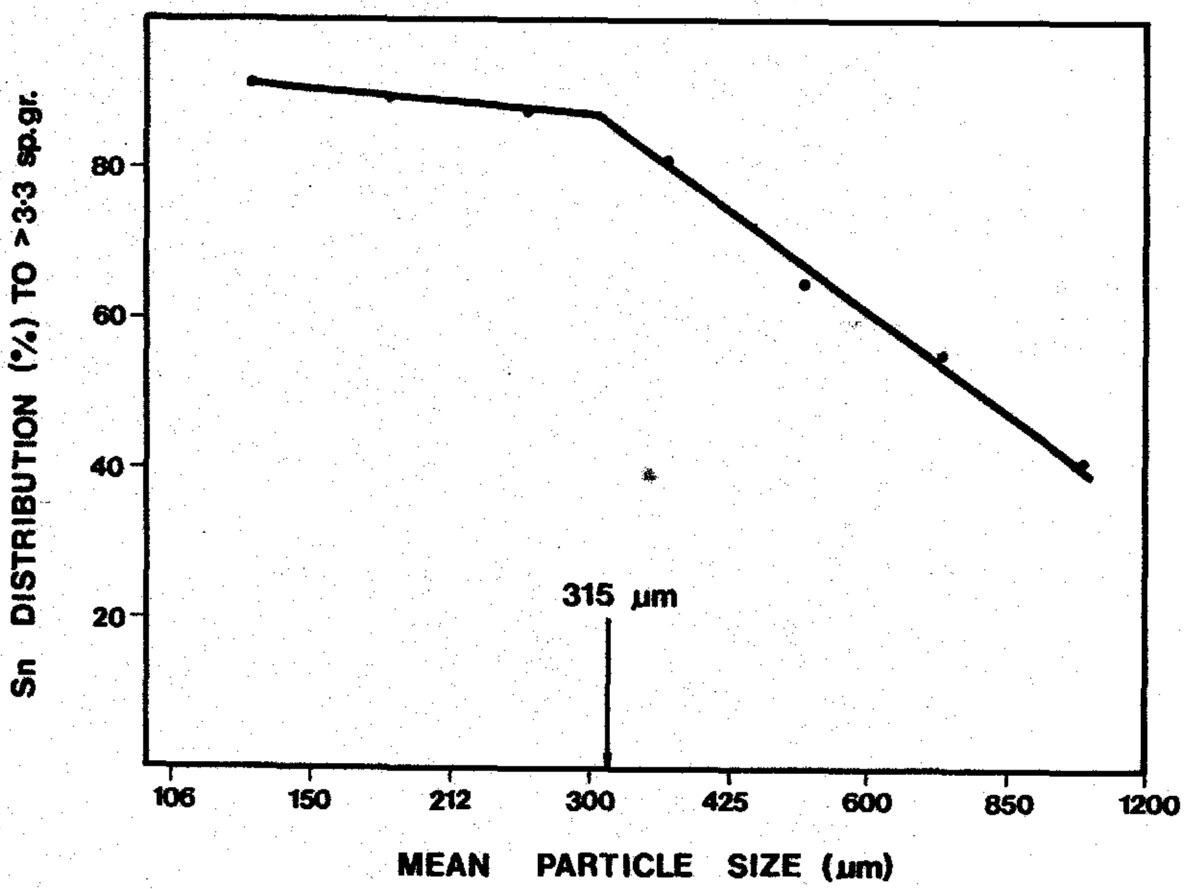

FIGURE AII-20

BT 59A

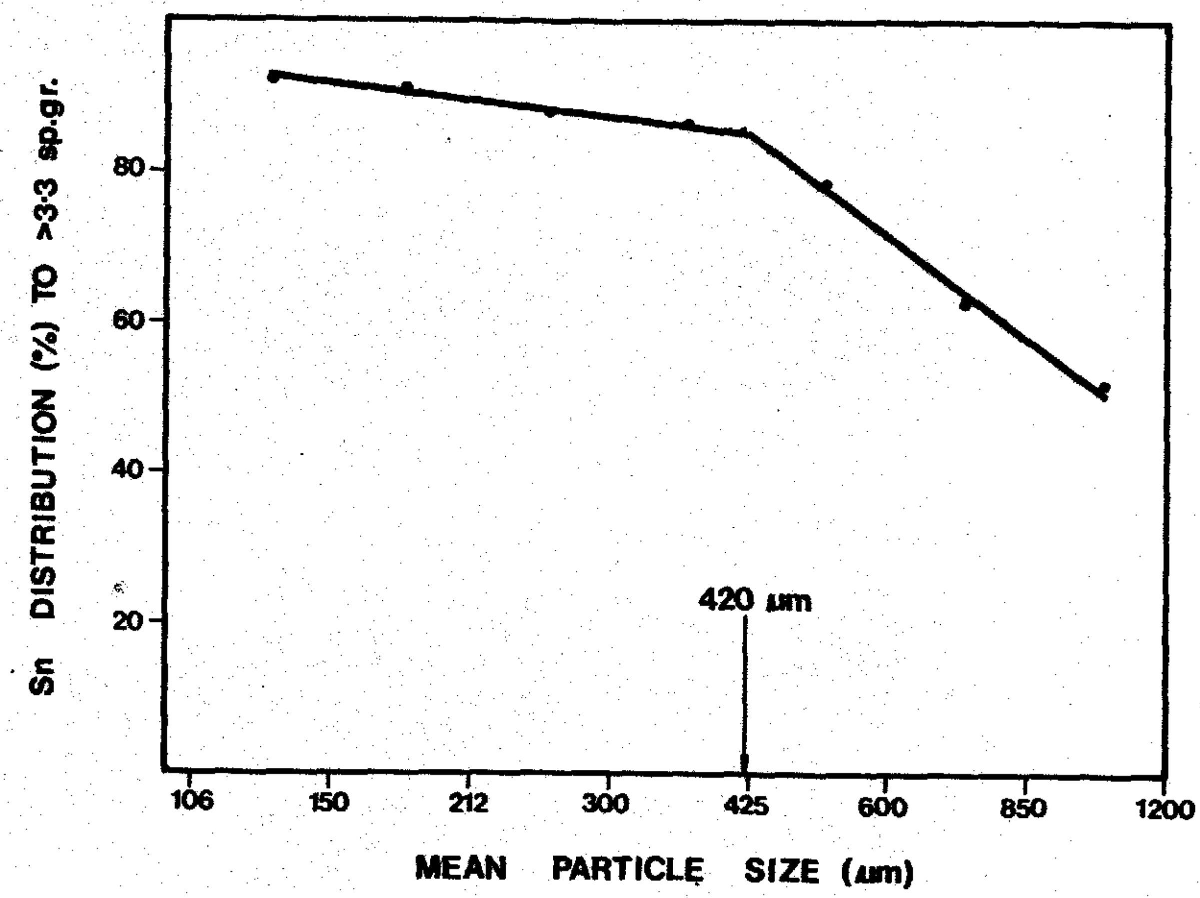

FIGURE AII-21

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084

BT 59B

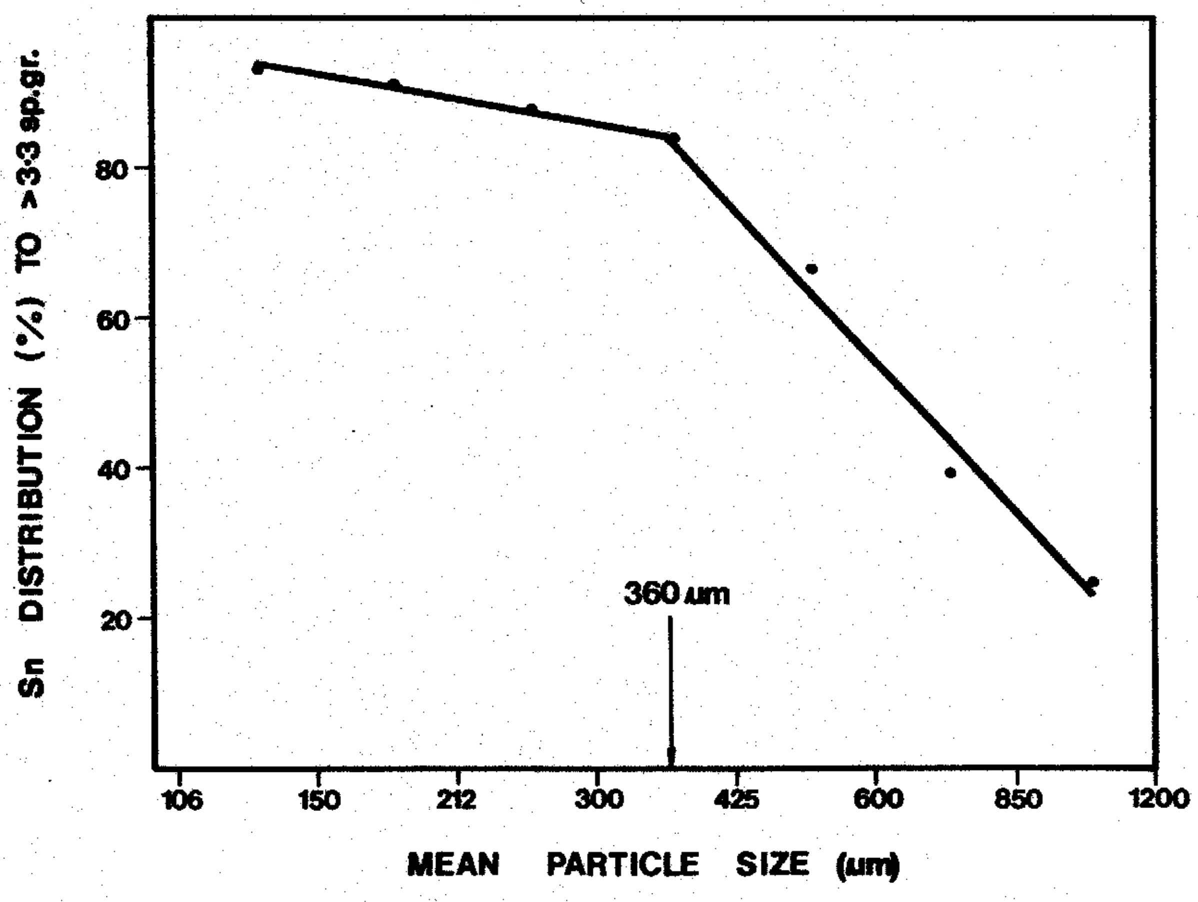

FIGURE AII-22

085

BT 69

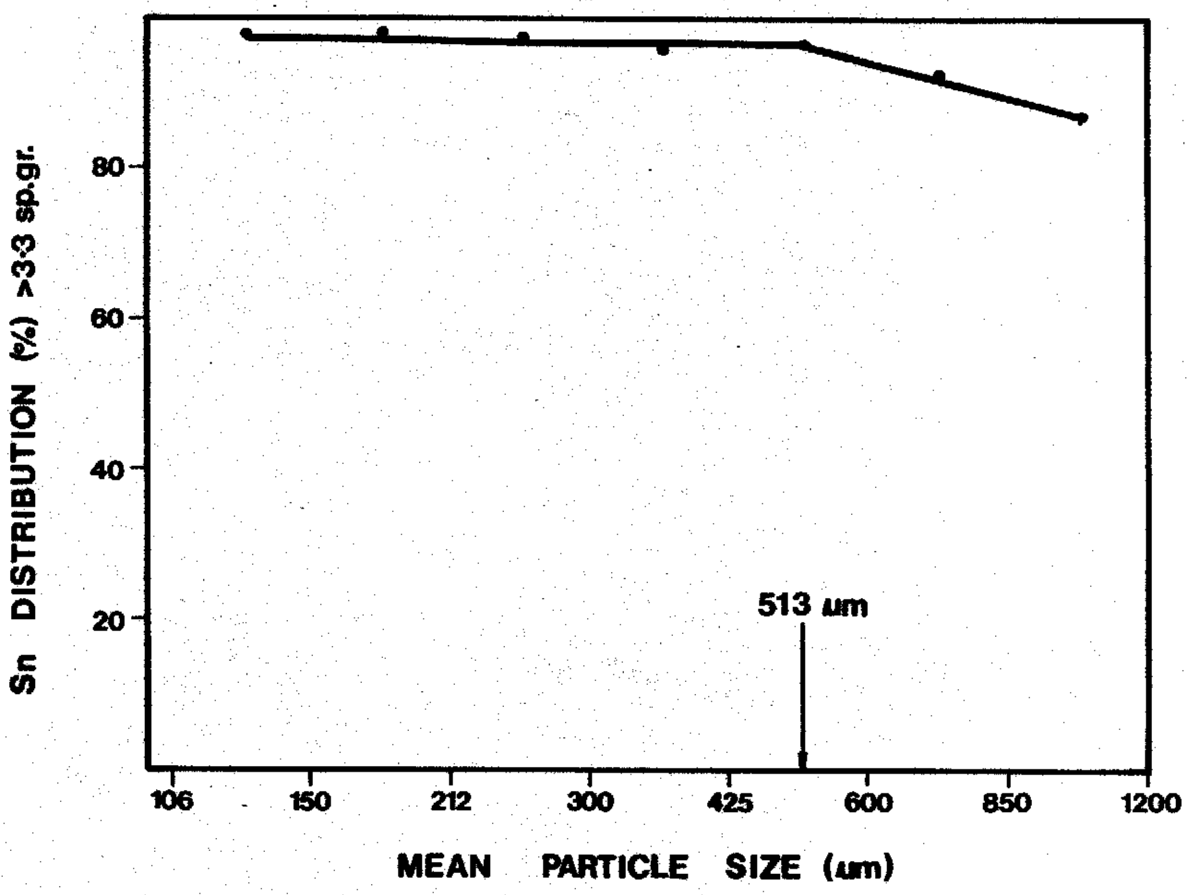

FIGURE AII-23

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086

BT 71

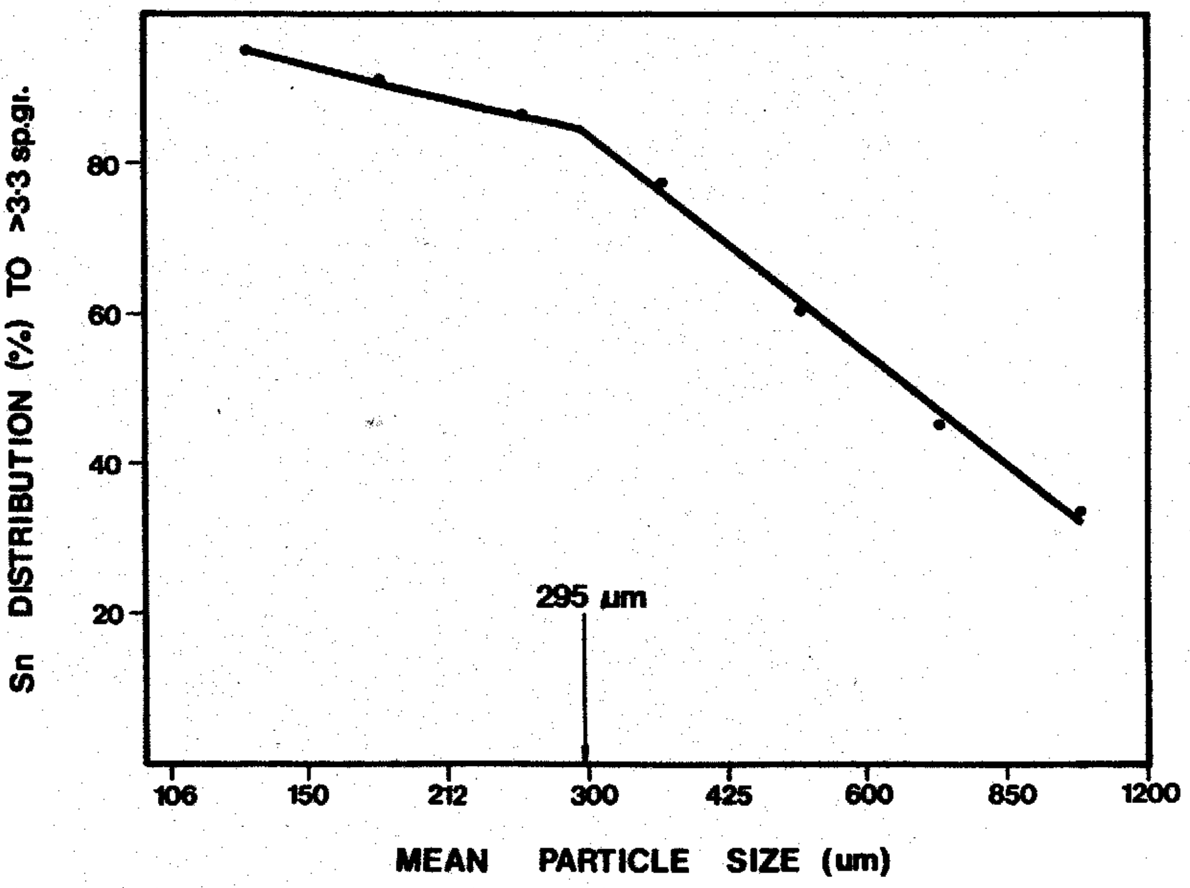

FIGURE AII-24

BT 81A

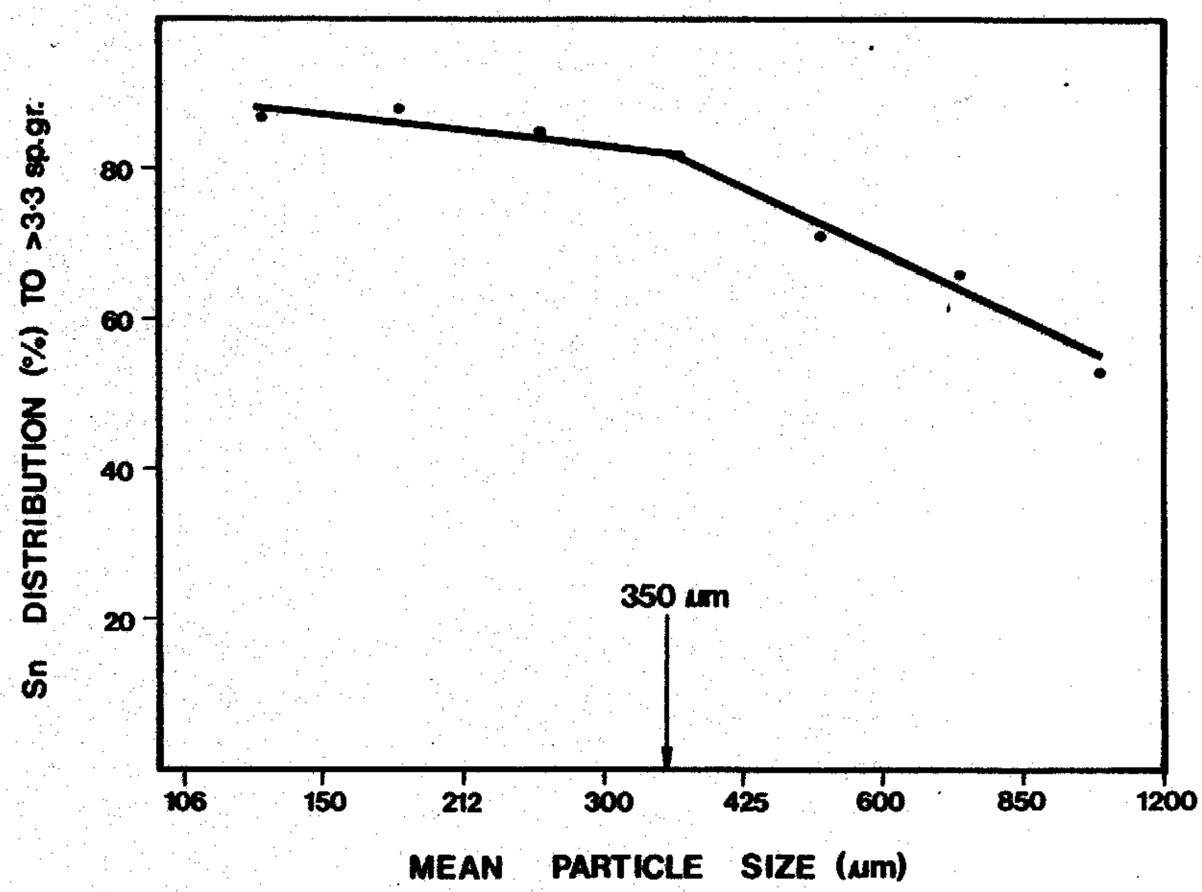

FIGURE AII-25

088

BT 81B

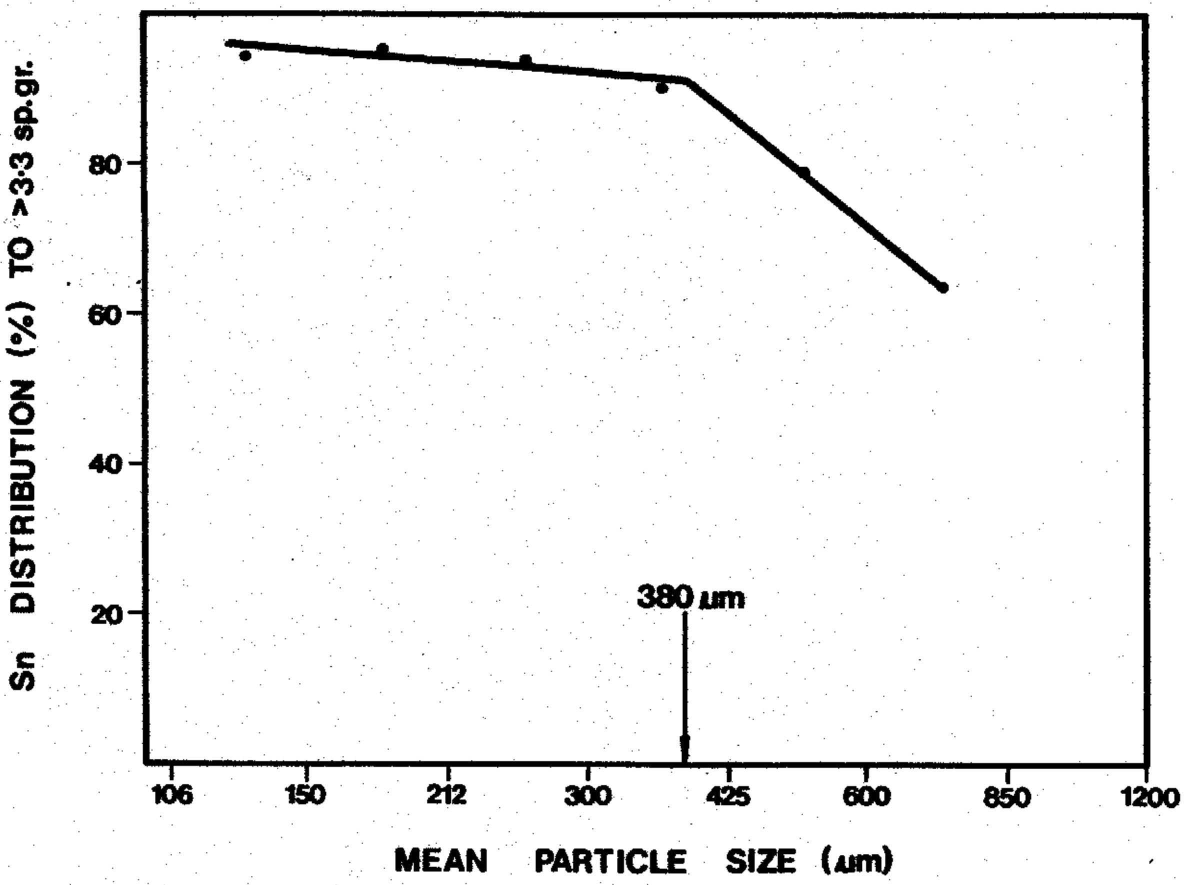

FIGURE AII-26

089

903092

BT 82

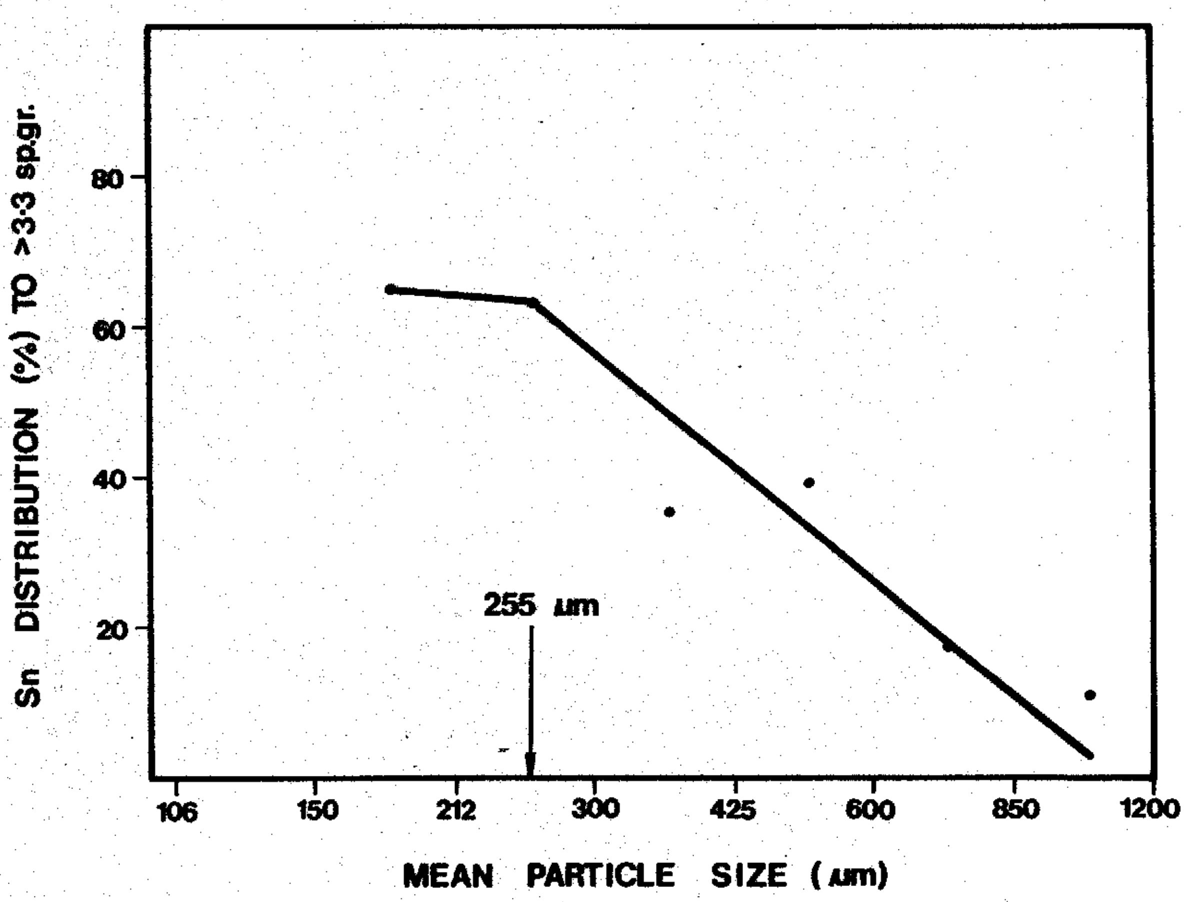

FIGURE AII-27

090

903093

BT 84

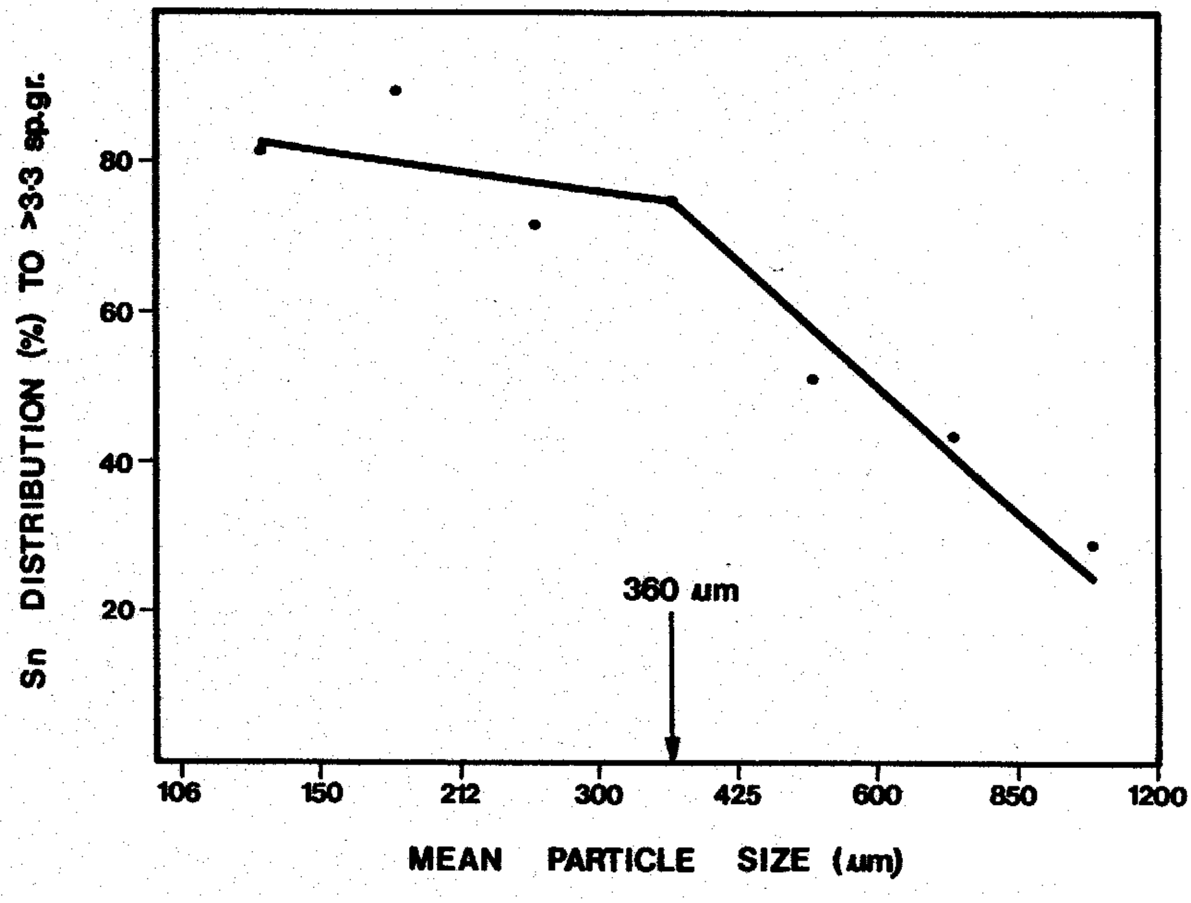

FIGURE AII-28

091

BT 86

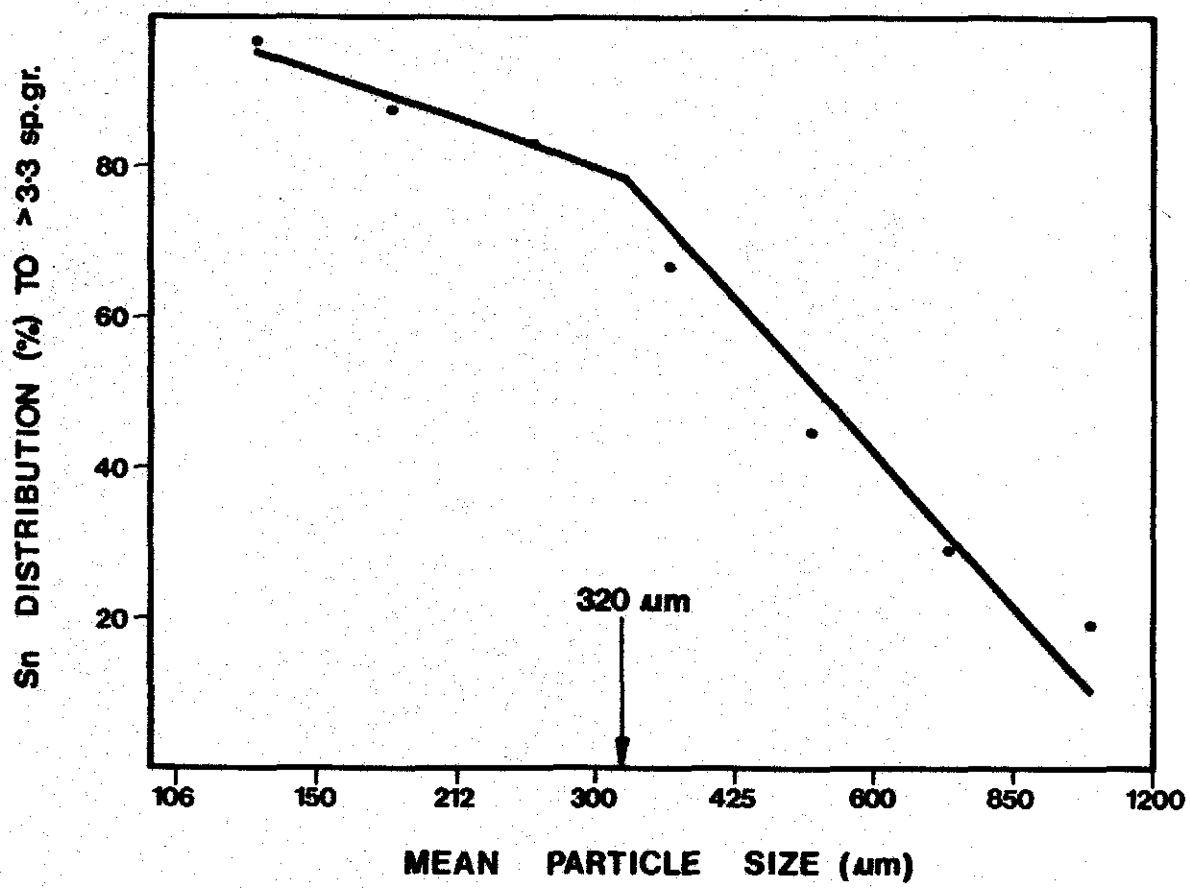

FIGURE AII-29

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092

BT 89

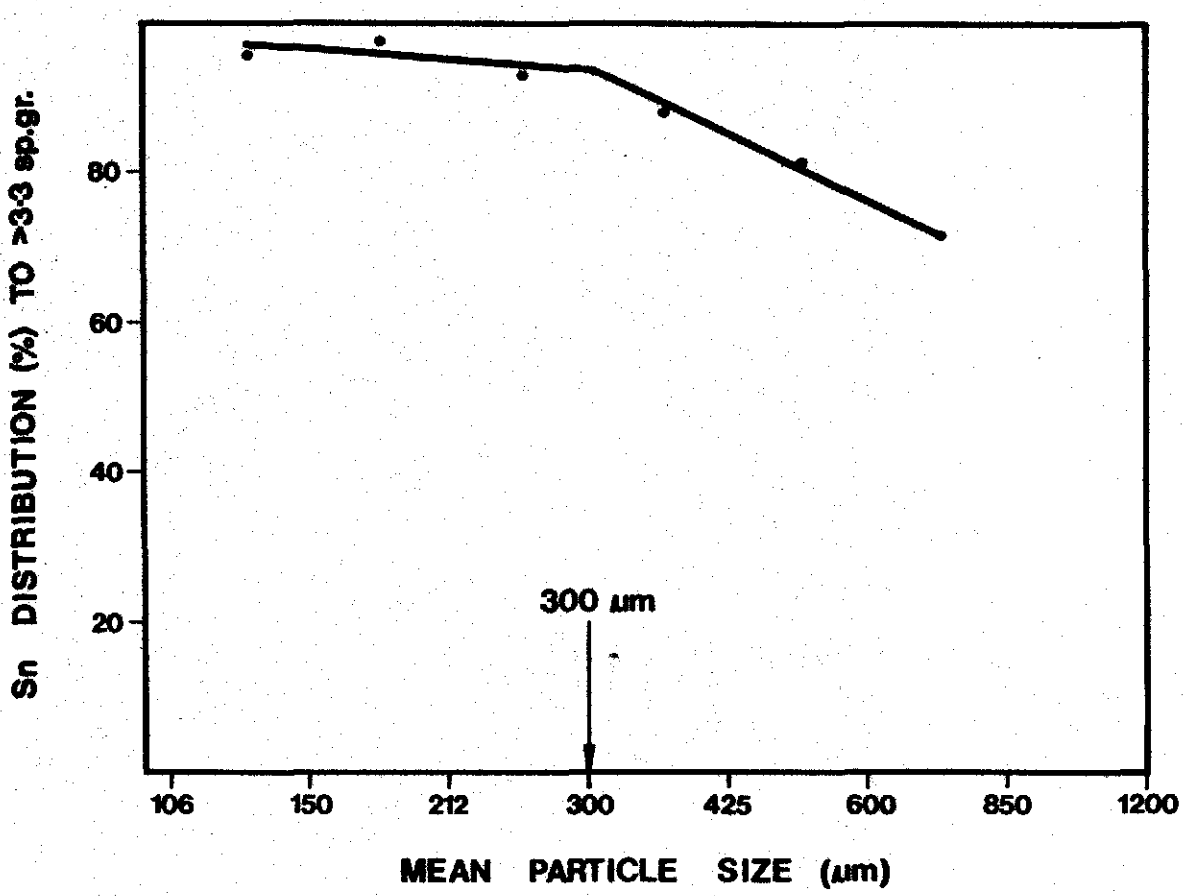

FIGURE AII-30

Table AII-1 Liberation Size

Sample	Size (μm)
Bulk Sample No.1	330
Bulk Sample No.2	215
Bulk Sample No.3A	300
Bulk Sample No.3B	190
BT 48	315
BT 59A	420
BT 59B	360
BT 69	560
BT 71	295
BT 81A	350
BT 81B	380
BT 82	255
BT 84	360
BT 86	320
BT 89	300

The sizes shown in Table AII-1 have a mean value of 330μm and a standard deviation of 88.

The required size of grind was estimated as 80 per cent finer than 350μm. This figure is slightly higher than the mean size calculated from Table AII-1 for several reasons.

1. Grinding produces fine material which is inefficiently recovered by gravity separation processes. Therefore maximising the grind size minimises minerals losses in the finest size fractions.

2. Maximising grind size minimises grinding capital and operating costs.

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3. Gravity separation testwork indicated that good recoveries to concentrates of acceptable grade were obtained even when the samples were ground coarser than the sizes indicated in Figures AII-15 to 19 (for example good grades and recoveries were produced from bulk sample No. 3B at a grind in which 80 per cent of the sample was finer than 400μm although a liberation size of 190μm is indicated in Figure AII-19.)

A consequence of not grinding sufficiently fine enough to achieve maximum liberation is reduced mineral recovery and concentrate grade. The example above indicates that recovery is not impaired by producing a coarser grind than that indicated.

In an operation of this scale mineral recovery would be expected to have a greater impact on revenue than concentrate grade.

AII-4 Classification

In order to achieve a grind in which 80 per cent of the product is finer than 350μm a screen with a 1mm aperture was selected.

AII-5 Variations in Grind Size

The variations in liberation size are neither of a sufficient magnitude nor of a consistent distribution within the orebody to warrant planned alterations in grind size. This is however a tool for tuning plant performance during operation.

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APPENDIX III

GRAVITY RECOVERY TESTWORK

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SAMPLE Nº 1

Key:

WT %	% Sn
% Sn DIST	

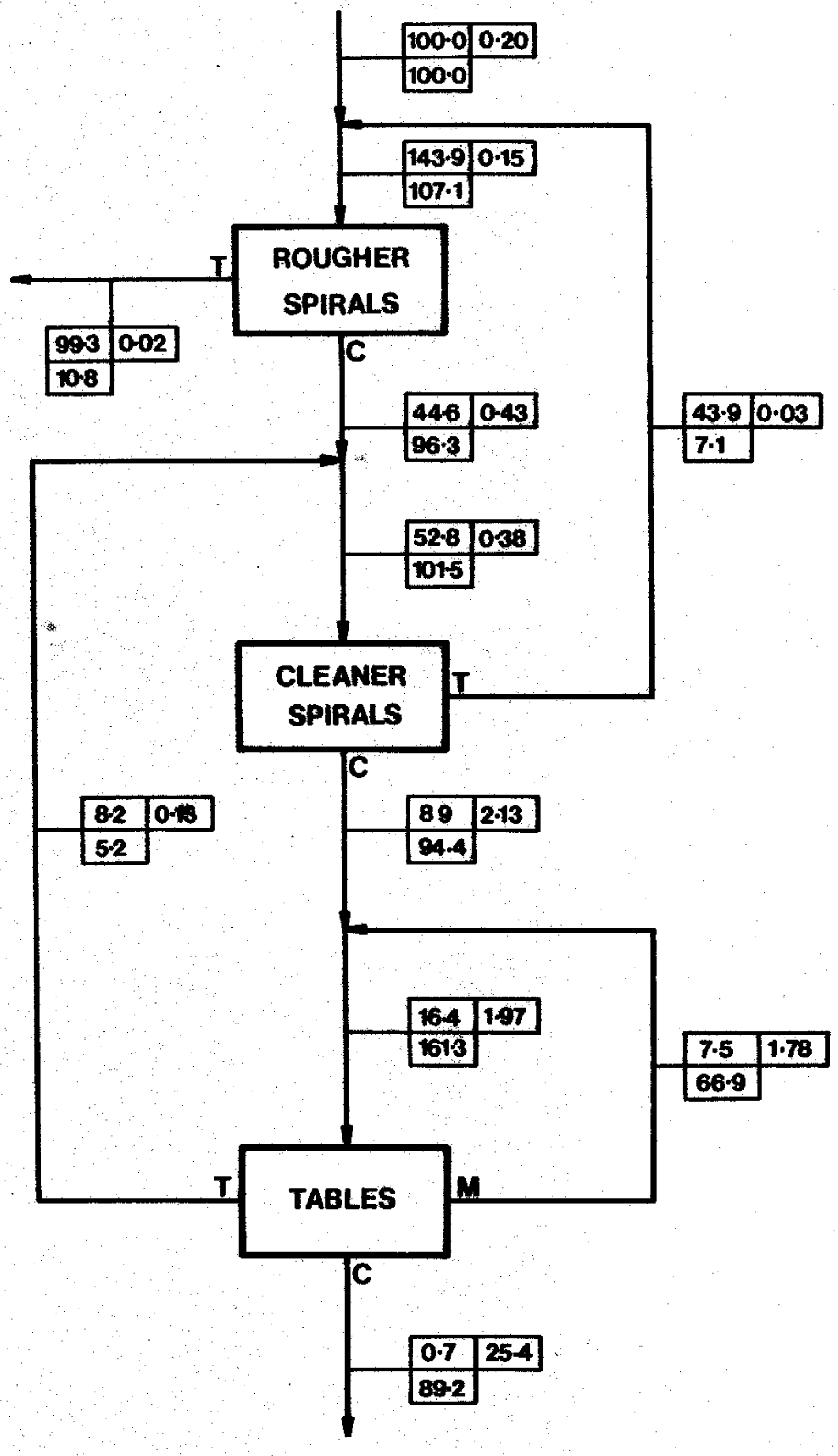


FIGURE AIII-1

Mineral Deposits Testwork

Initial tray and spiral testwork was carried out by Mineral Deposits on Bulk Sample No. 1.

Although results were extremely good for both the tray and spiral tests there were obviously some problems. Firstly, the calculated head grade from the spiral testwork was significantly lower than that obtained from initial subsampling and assaying of the bulk sample (0.20 per cent tin versus 0.35 per cent tin assayed). As several roughing tests were carried out in a closed circuit spiral test rig some loss of mineral due to settling in sumps or piping was indicated. This readily occurs in such tests if not anticipated.

Secondly the tabling results are mediocre. This can again be attributed to unfamiliarity with the material being treated and may have been compounded by a poor definition of test objectives. Nevertheless, encouraging results were obtained. A simple simulation model was used to estimate the effect of recycle streams on the overall performance of the circuit.

Results are summarised in Figure AIII-1. In the circuit shown in Figure AIII-1 middlings from the table are recycled to the feed to the table.

SAMPLE Nº 2

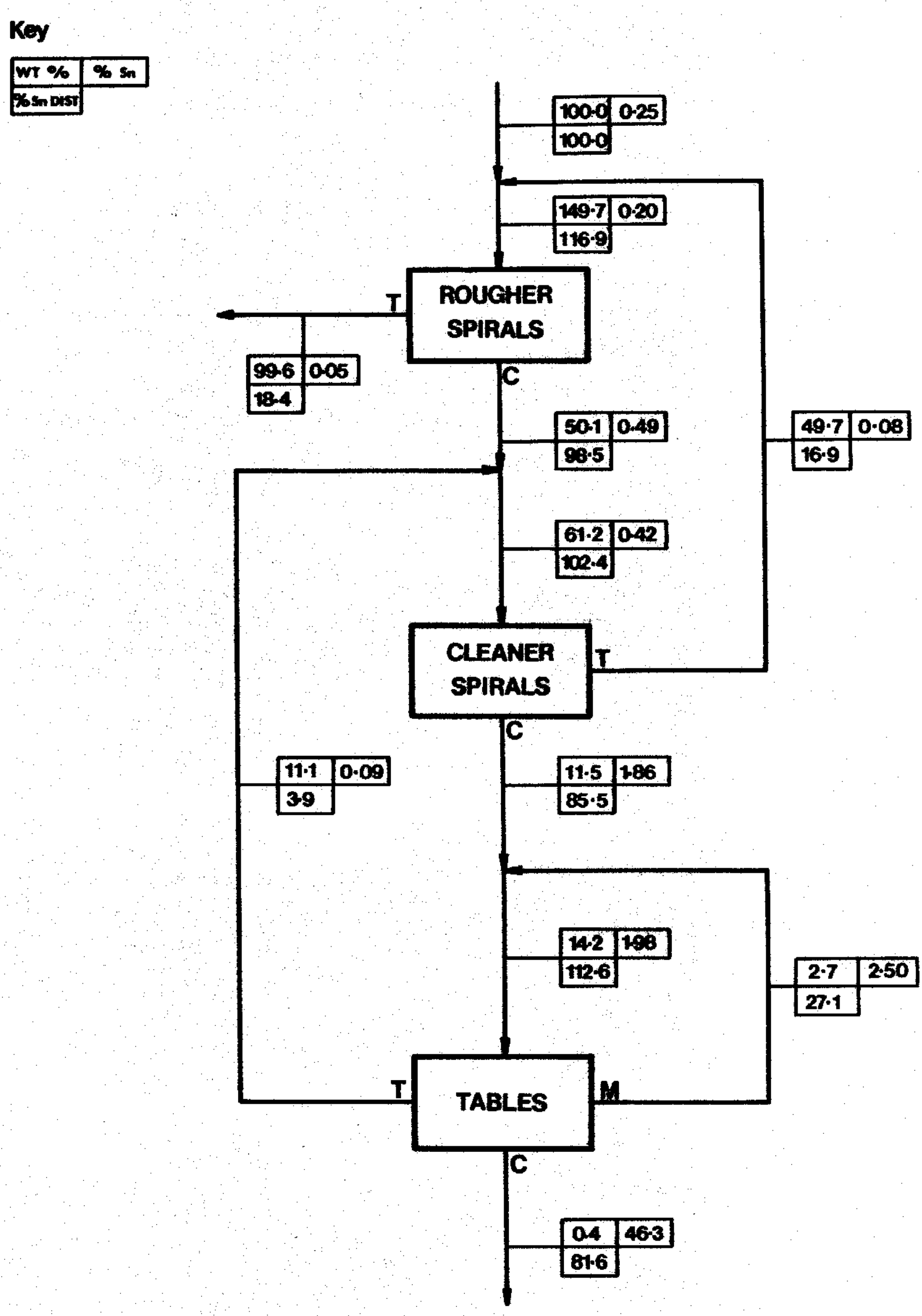


FIGURE A III-2

AIII-2 Testwork Carried out by Renison Personnel

Following the success of the Mineral Deposits testwork, tests using Bulk Samples No. 2 and No. 3B were carried out by Renison personnel using spirals manufactured by Vickers F.M.E.

Results were processed in a similar manner to those obtained from the Mineral Deposits tests and are summarised in Figures AIII-2 and AIII-3.

The spiral results obtained, especially for the roughing tests, are poorer than those for the Mineral Deposits tests. Size-by-size recovery data is shown in Figures AIII-4 and AIII-5 for the roughing and cleaning tests respectively. These results highlight the dramatic variations in performance obtainable from spirals of slightly different design. It becomes obvious that careful selection of equipment is essential to optimise performance and maximise return.

In this case the Mineral Deposits' spiral appears superior to the Vickers' spirals.

100

SAMPLE Nº 3B

Key

WT %	% Sn
% Sn DIST	

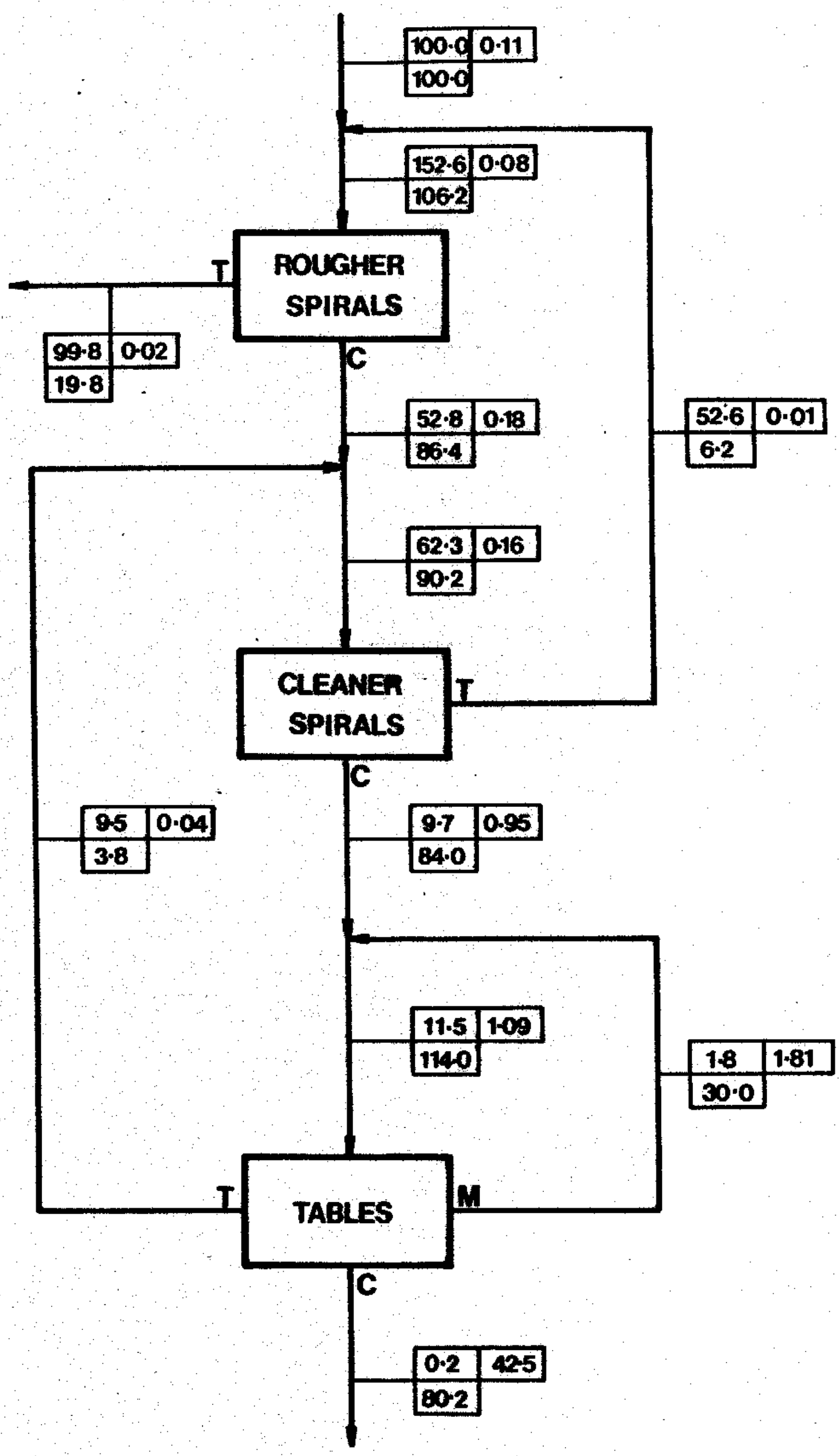
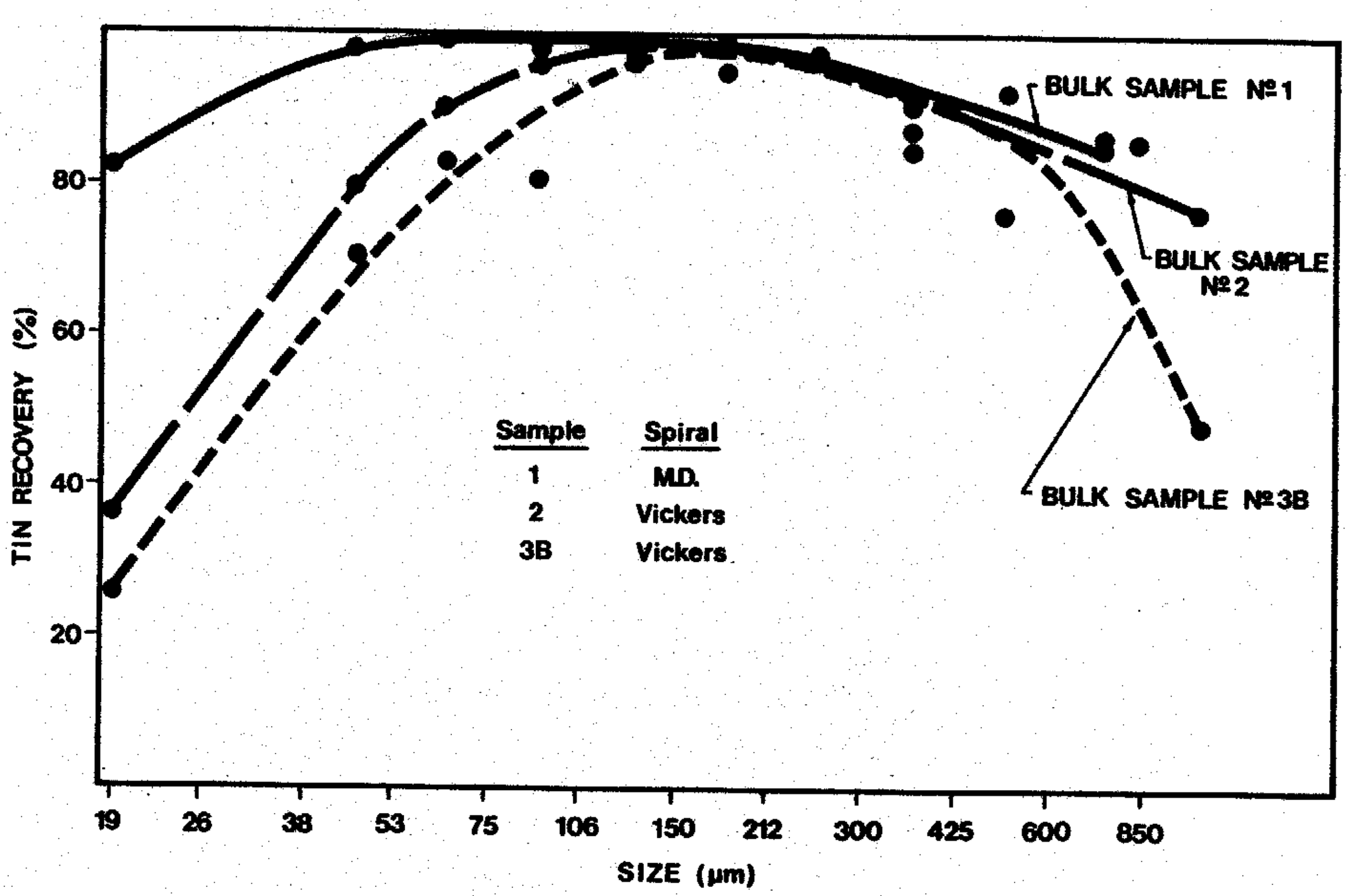


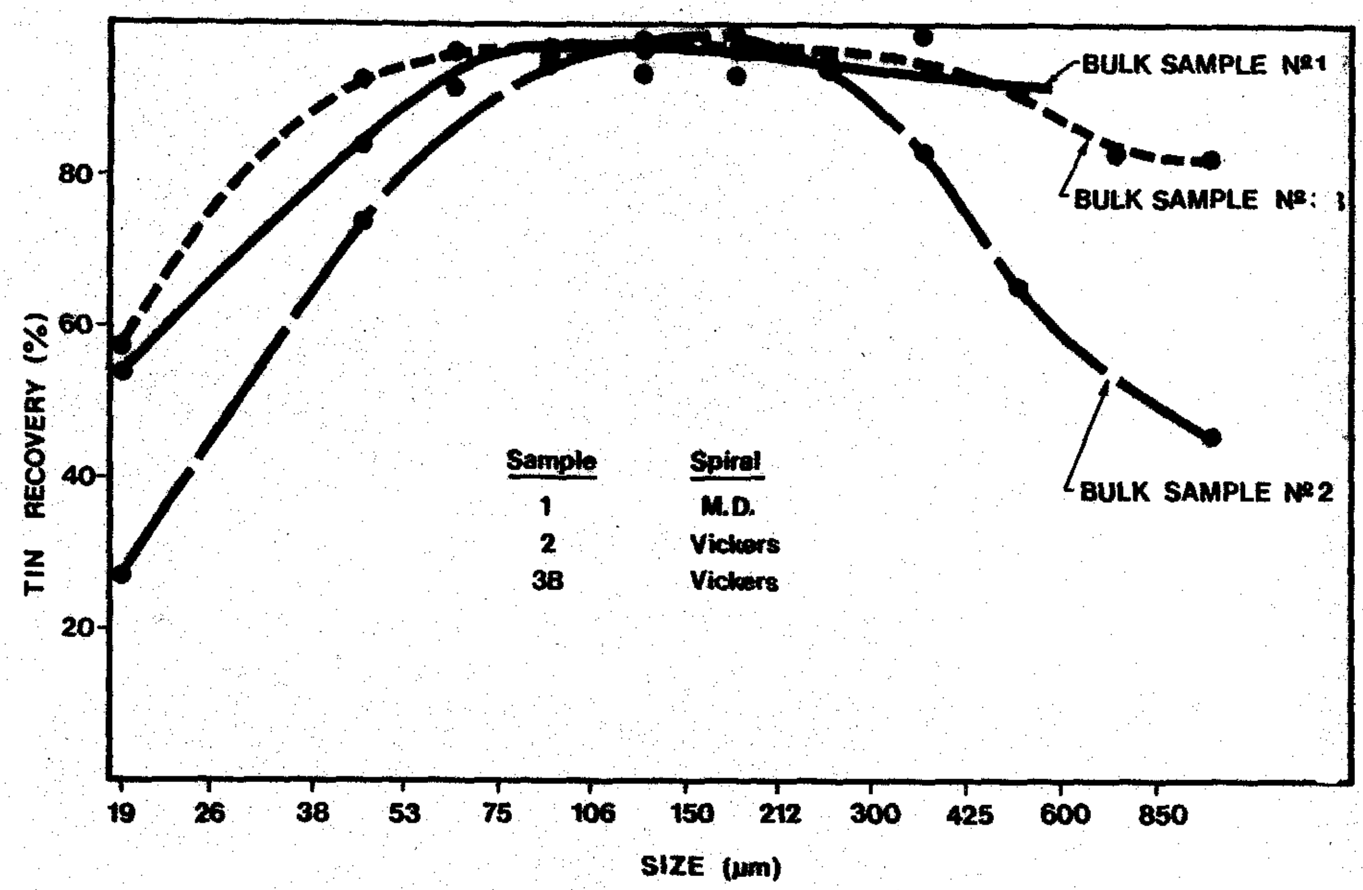
FIGURE AIII-3



SIZE BY SIZE RECOVERY FOR ROUGHER SPIRAL TESTS

FIGURE AIII-4

102



SIZE BY SIZE RECOVERY FOR CLEANER SPIRAL TESTS

FIGURE AIII-5

AIII-3 **Proposed Circuit**

Various circuit configurations were examined.

The proposed circuit incorporates retreatment tables for the further upgrading of tabling middlings rather than the recycle of table middlings to table feed. A comparison of Figures AIII-6 to AIII-8 with Figures AIII-1 to AIII-3 shows that significant improvements in concentrate grade with minimal reductions in recovery can be achieved on those obtained from the circuit initially envisaged. Such improvements require minimal additional tables.

AIII-4 **Prediction of Performance**

A simulation program was used to evaluate circuit performance when treating material of the same grade as the reserve. Stage recoveries for the Mineral Deposits' spirals were used in conjunction with the recoveries from tabling tests conducted at Renison to evaluate overall circuit performance when processing material with a 0.27 per cent tin grade.

Results are detailed in the main text (Section 4.5.7).

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SAMPLE Nº 1

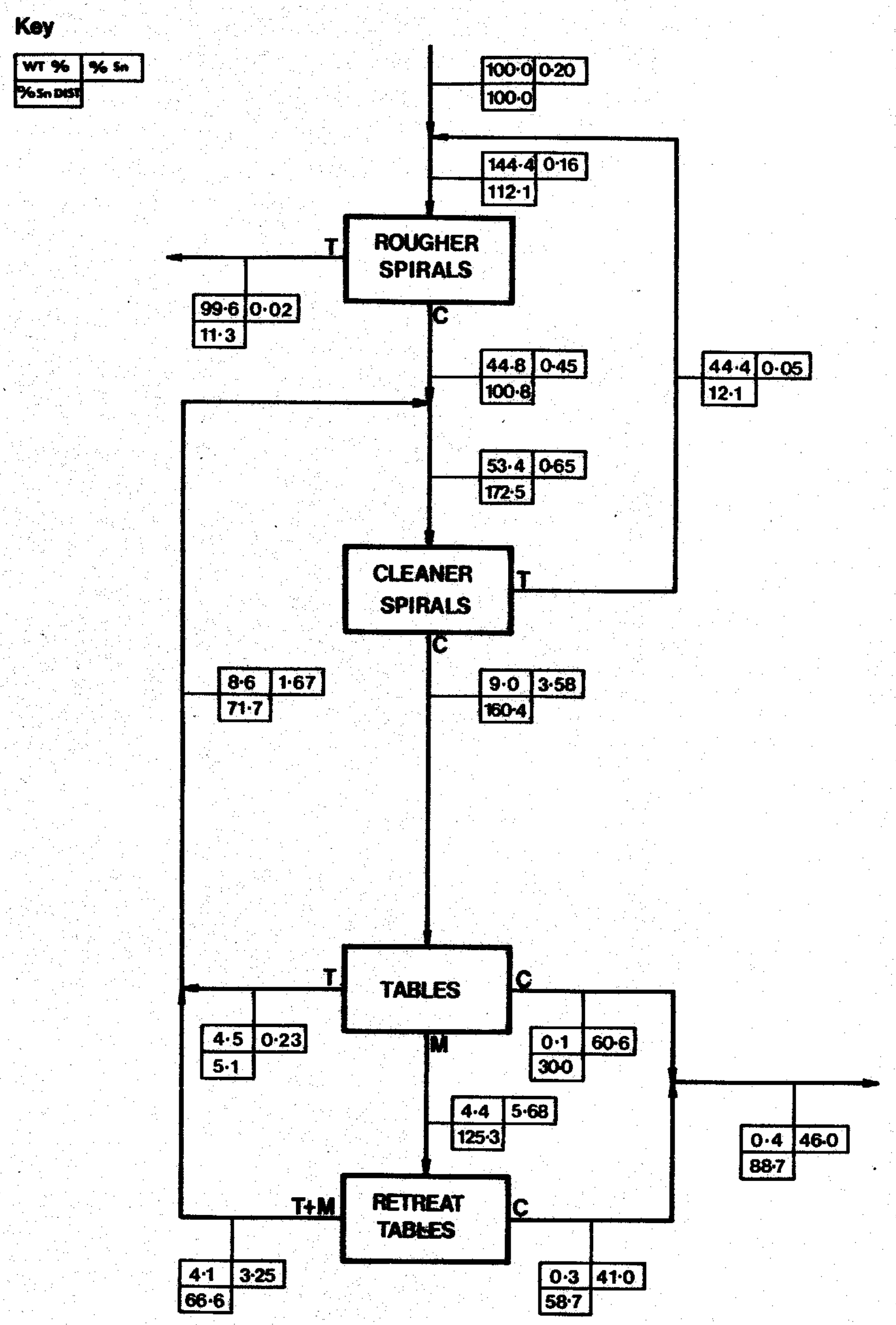

FIGURE AⅢ-6

SAMPLE Nº 2

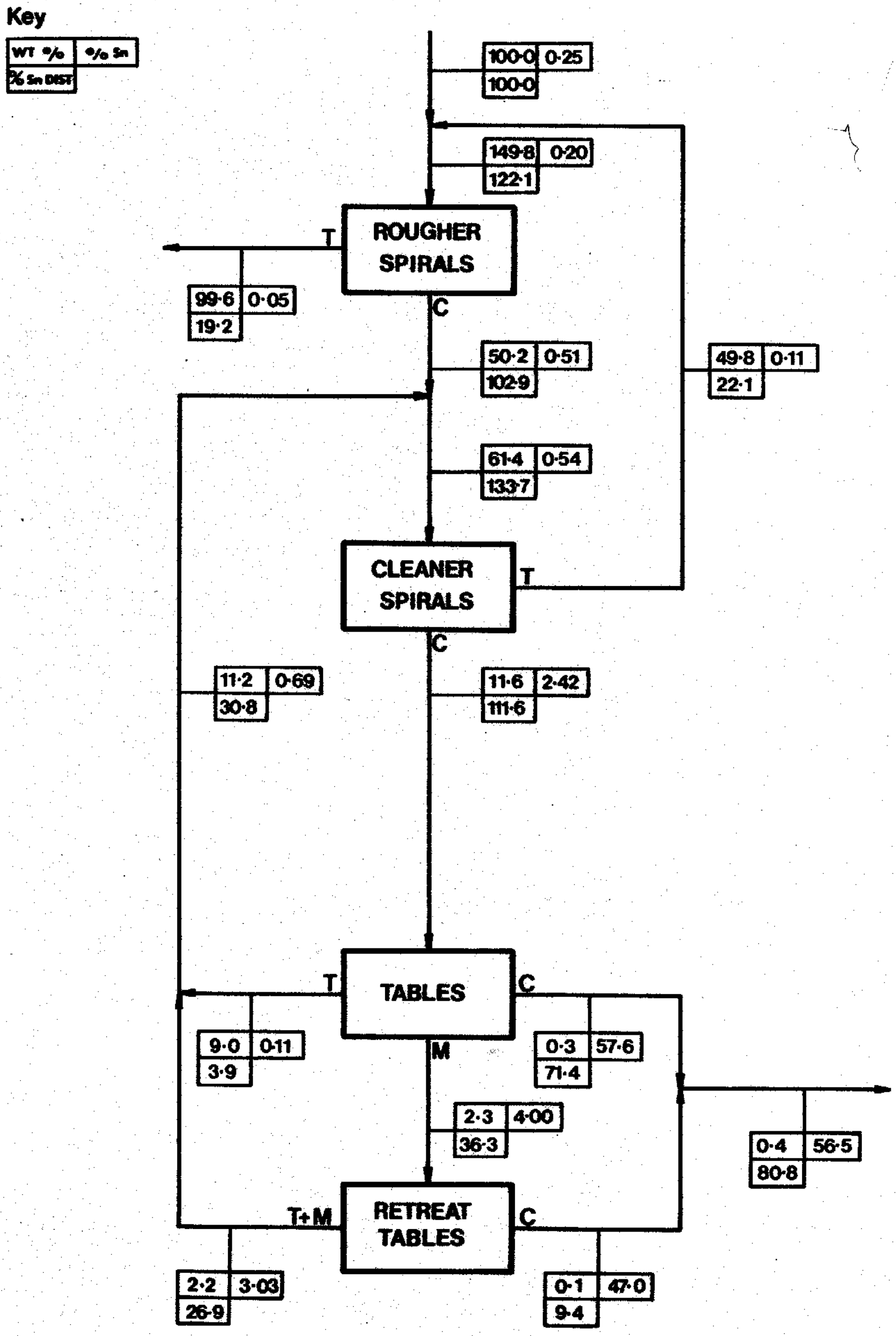

FIGURE AIII-7

SAMPLE Nº3B

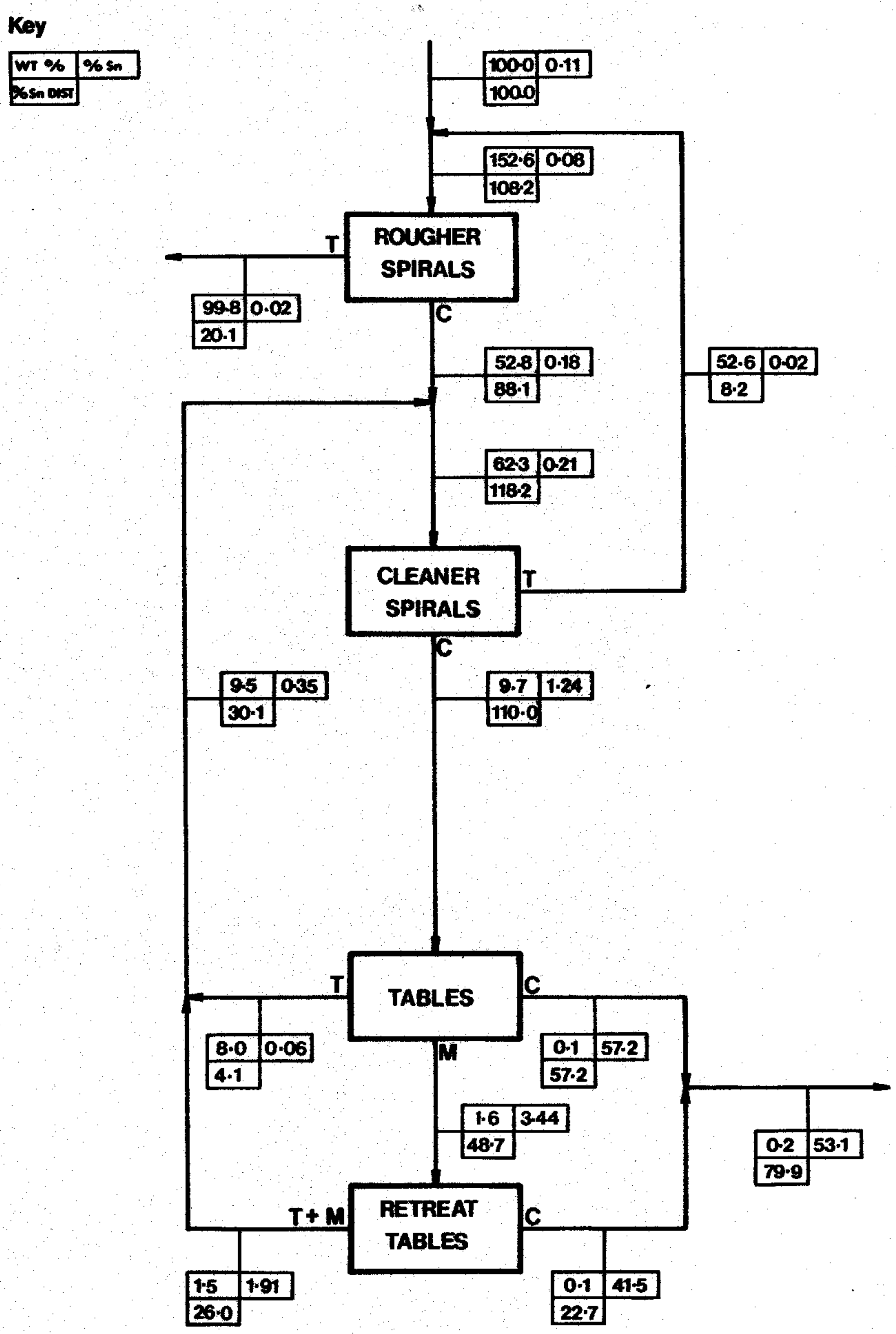

FIGURE AIII-8