uranium minerals can be obtained by some form of physical beneficiation before leaching.

4. ACKNOWLEDGEMENTS

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1. INTRODUCTION

In the early 1950s a worldwide uranium boom occurred, followed by an enthusiastic search for economic ore bodies in many parts of the world, including Australia. As a result of this activity a significant uranium occurrence was located at Mary Kathleen in July 1954. Extensive geological mapping, diamond drilling and metallurgical research followed. These activities proved that the ore body was economic and its development for production was commenced.

With a contract for the supply of yellow cake to the U.K.A.E.A. finalised in March 1956, construction of mine installations, the treatment plant and support facilities was commenced. The plant was in operation by June 1958 and continued production until 1963, during which time a total of 2,900,000 tons of ore were treated for a production of 9,000,000 lb of $\text{U}_3\text{O}_8$ as yellow cake. Since shutdown the plant and facilities have been maintained on a care and maintenance basis so that reopening can be achieved with the minimum of costly repair. Even though a large proportion of the treatment plant equipment and the buildings are exposed to the wind, rain and sun, the deterioration has been minimal, mainly as a result of the very dry air conditions which prevail nearly all year round.

Over the last few years there has been an increase in the use of nuclear power for electricity generation, and Mary Kathleen Uranium Ltd. has been able to secure contracts sufficient to justify reopening. On the basis of currently held sales contracts, it is planned to reopen the plant for production of yellow cake in late 1974. This will require the recommissioning of the original plant and installation of new equipment to begin at the end of this year.

It is envisaged that there will be a number of process changes and additions. The capacity of the plant will also be increased so that an output of two million pounds per year of $\text{U}_3\text{O}_8$ can be maintained with the future mill feed despite the fact that it will be of lower grade than during the previous operation. A number of process changes, discussion of which forms the basis of this paper, are also envisaged in order that operating costs can be reduced.

2. THE ORIGINAL FLOWSHEET

The existing treatment plant was based on a standard acid leach - fixed bed ion exchange recovery process. The flowsheet was not particularly unusual for its period although the concept of four stage continuous leaching with $\text{pH}$
and oxidation potential control was certainly advanced.

The basic process steps, which are illustrated in Figure 1, are as follows:

1. Primary crushing
2. Ore sorting
3. Secondary and tertiary crushing
4. Fine grinding
5. Leaching
6. Countercurrent decantation
7. Clarification
8. Ion exchange
9. Precipitation
10. Drying

As a number of these steps will be modified or replaced, it is useful to examine some of them as they existed previously so that the alterations can be more readily understood.

2.1 Primary Crushing and Sorting

Uranium bearing mineralisation is distinctly separate from barren minerals within the ore body. This permitted selective mining which upgraded the feed to the mill. The ore was trucked in 15 ton capacity rear dump trucks to the 48" x 60" Allis Chalmers primary jaw crusher. After crushing, the ore was screened at three inches, the oversize being conveyed to the ore sorter feed bins and the undersize to the surge bin ahead of the secondary and tertiary crushers. The plus three inch ore was fed to four radiometric ore sorters by vibrating grizzly feeders. As it left the feeder the ore dropped onto a rotating cone with a helical scroll. This caused the rocks to line up in single file for discharge onto a conveyor belt moving at about 180 feet per minute. As the rocks moved on the belt a mechanism positioned them in a straight line for final presentation to the discriminating apparatus.

When the particles reached the end of the belt they were discharged and fell past a light source/photo electric cell combination. From the variation in the intensity of the light, the area of the rock was measured. Still falling freely, the particle then passed a scintillation detector which measured the rock's radioactivity. The area and radioactivity signals were then compared. If the area signal was bigger than expected in relation to the radioactivity measurement for an ore particle of cut-off grade, the particle
was judged as waste. The electronic circuitry then activated three solenoid valves linked to air blast valves causing 90 psi compressed air to blast the waste particle aside at the correctly timed instant.

If the area signal indicated that the radioactivity measured was above the level for the cut-off grade set, the particle was allowed to fall freely onto the ore conveyor from which it was delivered with the minus three inch ore to the fine crushing surge bin.

### 2.2 Clarification

After the pregnant liquor overflowed from the No. 1 thickener in the countercurrent decantation circuit it was normally relatively clear (100 ppm solids). However, in order that the resin would not become contaminated, especially under surge conditions, when the pregnant liquor became excessively loaded with solids, the liquor was fed to a pair of rotary drum precoat filters, each with a surface area of 400 square feet. The precoat material was a mixture of two diatomaceous earth products. Occasionally the load was too great for the precoat filters and unclarified liquor was allowed into the ion exchange circuit without any apparent harmful effects.

### 2.3 Ion Exchange, Precipitation and Drying

After the pregnant liquor was clarified it was pumped to the ion exchange circuit consisting of three sets of four eight foot diameter columns, each loaded with 300 cubic feet of Deacidite FF resin, where absorption by ion exchange took place. Subsequently, a series of elution cycles were practised resulting in two concentrated uranium eluates relatively free of rare earths. (Rare earths were the contaminants of primary concern.) Elution was performed with return eluent made up to one molar sodium chloride and 0.1 normal sulphuric acid. This return eluent was recycled after precipitation of uranium from the normal eluate. The first major portion of the liquor produced during elution was separated and termed the high sulphate eluate. Since it was relatively low in chloride it was used as the sulphate bleed for the circuit after uranium precipitation. The second major portion, called the normal eluate, was higher in chloride, but lower in uranium and sulphate.

The high-sulphate eluate was neutralised to pH 7.0 with caustic magnesia (low in silica) to precipitate uranium. The yellow cake precipitate was thickened, filtered and washed on three successive rotary drum filters. The final filter cake was then dried at about 350°F in an oil fired conveyer type drier.

### 3. THE MODIFIED FLOWSHEET

Since the treatment plant operated at Mary Kathleen previously, a number of important developments have been tested and proven in other operating uranium mills and at Mary Kathleen as part of the development program. Where relevant, developments which would enable plant operating costs to be reduced have been incorporated into the proposed new flowsheet, which is shown in Figure 2.

#### 3.1 Ore Sorting

As has been already stated, sorting was carried out on the plus three inch ore after primary crushing. The maximum throughput was approximately 35 short tons per hour per sorter, with four sorters installed. Once the throughput was increased the sorting efficiency began to decrease.

Theoretically, the maximum capacity of a sorter installation is set by the inherent reaction frequency of the sensing head and attendant mechanical apparatus such as the blast valves. The maximum counting frequency under ideal conditions is believed to be six rocks per second.

With a maximum counting rate of six particles per second the spacing of particles on the belt from centre to centre for a belt speed of 180 ft/min would be six inches. For example five inch diameter particles would be spaced one inch apart and so on. During previous operation the optimum practical figure was about 4.5 rocks per second, taking into account the concentrate grade and recovery.

The need to depart from the ideal situation arises from two main factors.

1. A rock particle of high grade can influence the radioactivity reading for a waste particle thus resulting in its being accepted as ore. This situation is at its worst when a large ore particle precedes or follows a small waste particle.

2. The large range of particle sizes present in plus three inch ore can lead to a situation where the particle spacing becomes very difficult to control at an optimum figure (e.g. 7" rocks would overlap). The efficiency of blasting is affected
To lessen these problems, and at the same time increase the sorting installation capacity, it is planned to screen the ore sorter feed into at least two size ranges. Advantage will be taken of this increased capacity to present a slightly lower minimum feed size to the sorters. Therefore a greater overall beneficiation of the ore will be achieved with the same sorting equipment.

Whereas the maximum practical throughput for +3" ore as a single fraction was previously about 140 short tons per hour, for +3" ore split into two size fractions the throughput, as calculated from experimental test runs, should be increased by 15% for similar recoveries and grades. Initially, two screen sizes, -10" + 4 1/2" and -4 1/2" + 2 1/2", will be fed to the sorters. This will allow a balanced feed with two sorters being used for the coarser fraction and three sorters for the finer fraction. The fifth sorter will be a new one required in any event to handle the higher throughput needed to counteract decreasing feed grades.

The new fifth sorter will perform the same function as the others but its electronic circuitry will be built of solid state components as distinct from the older valve type construction. It is believed that this will lead to substantially better performance with lower maintenance costs. If this does occur, the other sorters will progressively be equipped with solid state components.

3.2 Sand Clarification

With a uranium solvent extraction process, the need for a clear pregnant feed liquor is greater than for ion exchange. A maximum solids content in the feed liquor of 20 ppm is expected to be acceptable. However, since on occasion during previous operation, the precoat filters were overloaded, the pregnant liquor throughput will be increased, and, the operating cost for the precoat filters was relatively high at nearly 3c/lb U₃O₈ produced, the possibility of replacing the precoat filters with pressure sand clarification columns was examined.

As explained below, the decision to replace the ion exchange plant with a solvent extraction system has been made. Thus it will be possible to convert some of the 12 ion exchange columns to pressure sand clarifiers.
economics of the alternatives of reusing the precoat filters, or installing pressure sand clarifiers using the existing ion exchange columns were compared by developing comparative cost figures. This demonstrated that pressure sand clarification would be superior.

Basically, pressure sand clarifiers can be operated on either the down-flow or the upflow principle. There are a number of different bed types available, apart from plain graded sand, which can take advantage of the different principles of operation. For example, if three materials of different specific gravity such as coal, garnet and silica sand are used in a bed then downflow operation can be used while minimising the harmful effects of clogging of the top bed layers as occurs with graded silica sand alone.

The final decision on the type of pressure sand clarifier has not been made but the leaning is towards the downflow type through a graded sand bed similar to that used at the Rum Jungle plant. Previous work at Amdel has also indicated that this method would be suitable for Mary Kathleen pregnant liquors.

3.3 Solvent Extraction, Precipitation and Calcination

When considering the choice between ion exchange and solvent extraction in a completely new application, a number of variations on these basic alternatives must be examined. If the resin-in-pulp process is applicable to the particular ore, then it is generally the cheapest route capital wise because there is a large saving as a result of the elimination of the solid-liquid separation circuit. However, solvent extraction generally results in a lower capital cost than fixed bed ion exchange. Relative operating costs often favour solvent extraction, but the degree of advantage is highly dependent on location.

The economic comparison between ion exchange and solvent extraction for Mary Kathleen was based on using the existing ion exchange plant with a new charge of a recently available tertiary amine resin. The original resin has deteriorated and lost too much of its loading capacity to be serviceable under full load conditions. This was compared with the alternative of modifying suitable equipment associated with the ion exchange plant to provide facilities for solvent extraction. Thus it should be emphasised that the comparison of solvent extraction in preference to ion exchange is based entirely on conditions peculiar to Mary Kathleen.

The evaluation showed that in spite of the higher capital expenditure necessary for the solvent extraction plant, the reduction in annual operating costs more than compensated for this and justifies the installation of a solvent extraction plant in the Mary Kathleen situation. The proposed solvent extraction method, as shown in Figure 3, is basically the Purlex process. It involves extraction of the uranium from the clarified pregnant liquor using a tertiary amine solvent in a kerosene diluent. A small percentage (3%) of a fatty alcohol (nonanol) is also included in the kerosene to help with phase disengagement.

After countercurrent extraction in four stages, the solvent will be scrubbed with water in a single stage to reduce the level of impurities such as rare earths. Stripping will then take place in four stages using ammonium sulphate, with pH control being maintained on stages 2 and 3 using 5% aqueous ammonia solution. Intermittent scrubbing of the solvent with sodium carbonate will also be performed. The loaded aqueous strip solution will finally be adjusted to pH 7.5 in three stages with ammonia to precipitate uranium. The yellow cake precipitate will then be thickened to about 30% solids and pumped to a centrifuge where a cake of about 60–70% solids will be produced. This cake will be calcined at 500°C in a multiple hearth drier.

A number of the decisions related to the selection of various alternatives in this final section of the treatment plant flowsheet are integrally related. Ammonia was chosen for stripping pH adjustment and for yellow cake precipitation because it is basically cheaper than other alkali reagents for these operations, and it also enables a very high grade product to be produced by calcination, thus reducing the cost of transport per unit of contained U. Ammonia is driven off in the calcination step whereas the sodium and the magnesium in their respective diuranate yellow cake precipitates remain after calcination. In addition, it is worth noting that, because of the geographical location of Mary Kathleen, an important factor in the selection of ammonia as a more economical reagent than caustic soda or calcined magnesia, is its low molecular weight and hence the reduced freight costs associated with using it.

A comparative list of operating requirements and costs is shown in Table 1 for:

(i) an ion exchange process based on previous experience at Mary Kathleen and recent experimental results with modern weak basic tertiary amine resins, and
(ii) the proposed solvent extraction plant. Precipitation is also allowed for in these requirements.

The effect of the high cost of transporting raw materials to Mary Kathleen has a strong influence on the economics of ion exchange, which uses a larger weight of raw materials in process. Weight can be more important at Mary Kathleen than basic chemical cost. Sodium chloride is an extreme example of this situation. It is also expected that some labour reduction will be possible with the solvent extraction plant.

A solid bowl scroll discharge type centrifuge has been chosen as the replacement for the three stage filtration system for the final yellow cake product. Though the economics are only marginally in favour of the centrifuge there are other qualitative factors which make the choice decisive. One important feature is the elimination of possible contamination of the working area by the enclosed operation of the centrifuge. Another is its ability to handle a slimy precipitate.

3.4 Additional Modifications

As well as the alternations already described several other changes to the circuit will be made. The countercurrent decantation circuit (CCD) consisting of five 75\textdegree\ diameter thickeners was previously close to capacity during the final months of operation. Since the ore throughput will increase it is anticipated that the CCD circuit could become overloaded. Therefore a cyclone will be installed to scalp off the heavier and coarser particles of leached ore and only allow the finer and lighter material to reach the thickeners. Coarser material will be washed in a further series of cyclones.

In keeping with the policy of eliminating as much waste as possible, wash water for the CCD circuit will be made up partly by recycle from the liquor dam. At the same time some soluble uranium will be recovered which would otherwise be lost. This raffinate, which has a pH around 2.0, will also be utilised in the heap leaching of waste and low grade ore before recycle, producing some additional uranium and permitting acid to be recovered from the raffinate.

4. CONCLUSION

We hope that this brief discussion of the bases for proposed changes in the Mary Kathleen flowsheet will help to illustrate the need for a unique approach to any new uranium processing requirement and the need for constant review of the situation even after the initial choice is made.
### TABLE 1

**COMPARATIVE OPERATING REQUIREMENTS AND COSTS**

<table>
<thead>
<tr>
<th></th>
<th>Solvent Extraction</th>
<th>Ion Exchange</th>
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<tbody>
<tr>
<td></td>
<td>Usage</td>
<td>Cost</td>
</tr>
<tr>
<td></td>
<td>lb/lb U₂₃₀</td>
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<tr>
<td><strong>Raw Materials</strong></td>
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<tr>
<td>Kerosene</td>
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<tr>
<td>Amine</td>
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<tr>
<td>Nonanol</td>
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<td>1.46</td>
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<tr>
<td>Sulphuric Acid</td>
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<td></td>
</tr>
<tr>
<td>Sodium Chloride</td>
<td></td>
<td></td>
</tr>
<tr>
<td>Caustic Soda</td>
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<td></td>
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<tr>
<td>Ammonia</td>
<td>0.435</td>
<td>2.96</td>
</tr>
<tr>
<td>Resin</td>
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<tr>
<td><strong>Sub Total</strong></td>
<td>5.43</td>
<td></td>
</tr>
<tr>
<td><strong>Labour</strong></td>
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<tr>
<td>2 men/shift + 1 man day shift</td>
<td>2.57</td>
<td>3 men/shift 2 men day shift</td>
</tr>
<tr>
<td><strong>Operating Supplies</strong></td>
<td>10% of labour</td>
<td>0.26</td>
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<td><strong>Utilities</strong></td>
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<td>Water</td>
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<tr>
<td>Electricity</td>
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<tr>
<td><strong>GRAND TOTAL</strong></td>
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</table>

**NOTE:** Costs are in 1972 terms.