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AUSTRALIAN ATOMIC ENERGY COMMISSION
RESEARCH ESTABLISHMENT
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December 1978

ISBN 0 642 59674 3

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ABSTRACT

Developments in uranium ore processing technology since the AAEC Symposium on Uranium Processing in July 1972 are reviewed. The main developments include the use of

- autogenous or semi-autogenous grinding,
- beneficiation techniques such as radiometric sorting, flotation, magnetic and gravity separation,
- strong acid and ferric bacterial leaching processes,
- solution mining and heap leaching operations,
- horizontal belt filters for solid-liquid separation,
- continuous ion exchange processes for use with solutions containing up to 8 wt % solids,
- hydrogen peroxide and ammonia for the precipitation of uranium to improve product yield and purity, and
- the recovery of by-product uranium from the manufacture of phosphoric acid and copper processing operations.

National Library of Australia card number and ISBN 0 642 59674 3

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URANIUM; URANIUM ORES; ORE PROCESSING; CRUSHING; RADIOMETRIC SORTING; FLOTATION; LEACHING; SOLUTION MINING; SOLVENT EXTRACTION; ION EXCHANGE; CHEMICAL REACTION YIELD; PURIFICATION; HYDROGEN PEROXIDE; AMMONIA

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1. INTRODUCTION

This report summarises developments in the technology for uranium ore processing which have occurred since the AAEC Symposium on Uranium Processing held in July 1972 [AAEC 1972]. The status of the technology had not long before then been reviewed at various conferences [e.g. IAEA 1971] and in a number of publications [e.g. Merritt 1971].

In the intervening period, the projected demand and price of uranium [OECD/IAEA 1977] has stimulated interest and research in uranium recovery. There have also been significant increases in capital and operating costs, caused largely by the processing of lower grade ores and inflationary pressures, but also by increases in the cost of reducing the environmental impact of mining and milling. In general, rising costs have provided an incentive to improve the conventional extraction methods. Higher uranium prices have also increased interest in extraction of uranium from lower grade ores, for example by *in situ* leaching, and by the commercialisation of processes developed for recovery of uranium as a by-product of phosphoric acid and copper processing operations.

Many of these developments have been reported at recent major conferences [IAEA 1976a; IAEA 1977; IMM/CEC 1977]. In particular, the IAEA [1976a] provided a comprehensive summary of the commercial flowsheets used in past and present uranium mills and proposed for new plants or deposits under investigation, as well as a summary of existing, improved and new techniques for uranium ore processing steps*.

2. PROCESS ROUTES

The basic steps in the conventional processes using acid and alkaline leaching are shown in Figures 1 and 2, respectively. Some low grade ore may be rejected by various beneficiation techniques before

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Footnote: *Following the preparation of the present review, H.E. James and H.A. Simonsen presented a paper on 'The Significance of Ore Processing Technology in the Assessment of the Uranium Supply Outlook' to the International Symposium on Uranium Supply and Demand, sponsored by the Uranium Institute, London, 10-12 July 1978. This paper comprehensively reviewed the nature of the major types of uranium ore deposits, the effects of their mineralogy on processing technology and of improvements in this technology on uranium resources and supply.*

leaching. The choice of leachant depends on the chemical composition of the ore, particularly the presence of acid-consuming gangue minerals. About 80 per cent of the world's uranium is extracted by leaching with sulphuric acid to which an oxidant, commonly pyrolusite or sodium chlorate, is added to convert the tetravalent uranium to the more soluble, hexavalent uranyl species. After leaching, the tailings are washed and separated from the leach liquor by multistage countercurrent decantation or filtration. The uranium in the leach liquor is purified by solvent extraction or ion exchange, precipitated, separated, dried and calcined to produce yellowcake.

The alkaline leaching route using sodium carbonate-bicarbonate solution follows similar steps but because the leaching stage is more selective for uranium it is not necessary to purify the leach liquor before the uranium is precipitated with caustic soda as sodium uranate. This material may be dissolved in sulphuric acid and re-precipitated with ammonia to give a higher grade product. The caustic barren liquor is regenerated with carbon dioxide to restore the carbonate-bicarbonate balance and recycled for further leaching.

3. CRUSHING AND GRINDING

Typically, primary crushing (to 50-150 mm) is carried out in large jaw crushers, with secondary crushing (to 10-50 mm) in smaller jaw or gyratory crushers and fine grinding in rod and ball mills to 30-50 per cent -200 BSS. Wet grinding has been widely practised in the past in Australia, USA, Canada, South Africa and France [IAEA 1976a]. Pebble mills have also been widely used in both South Africa [IAEA 1976a] and Canada [IAEA 1976a; Lendrum 1974; Lendrum & McCreedy 1976]. In Canada, rod plus pebble milling is used at the Denison mill [Murphy 1975a] and at the Rio Algom Quirke mill [Murphy 1975b]. The use of pebble mills reduces grinding steel requirements and, by reducing the amount of iron in the leaching circuit, also reduces acid and oxidant consumption. It has been found in Canada that less wear results if the pebbles are taken from the primary jaw crusher rather than from the secondary crushing stage [Lendrum & McCreedy 1976].

Until the early 1970s, autogenous grinding was used at only two mills, at Eldorado's Beaverlodge mill in Canada [Fish 1975; Lendrum & McCreedy 1976] and in Union Carbide's Uravan mill in Colorado [Merritt 1971]. Since then, autogenous or semi-autogenous grinding has been incorporated into a number of new uranium mills and proposed for further mills (Table 1). In autogenous milling, the ore acts as its own grinding medium whereas in semi-autogenous milling a small proportion of steel balls is added. Depending on the ore, wet semi-autogenous mills can accept ore up to 0.6 m in diameter and grind both large and small rocks to the feed size required for leaching [Jones 1977]. The crushing section of the mill is thus potentially eliminated (although a primary crushing stage may be included) and problems from sticky ore, dust and noise are greatly reduced. The method is well suited to new operations which must meet rigorous standards for environmental and health protection [McManus 1977].

Semi-autogenous grinding may be more economic; at Gulf's Rabbit Lake mill, the investment cost for the semi-autogenous circuit was about 40 per cent of the cost for conventional grinding, the power installed was reduced by 50 per cent and the wear during operation by 40-50 per cent [Flöter 1976].

4. BENEFICIATION

Pre-concentration methods have been investigated for over 20 years, but there have been only limited commercial applications, except in South Africa where flotation is widely used for concentration of uranium, gold and pyrite from Witwatersrand ore. However, recently there has been renewed interest in beneficiation as a result of the incentive to process lower grade ores. At the same time, past experience indicates that the probability of achieving both a high recovery and a high concentration ratio is low for most uranium ores.

4.1 Ore Sorting

4.1.1 Radiometric sorting

Radiometric sorting has been widely used in the USSR [IAEA 1976a; Skorovarov *et al*. 1976; Skrinichenko *et al*. 1977]. Applications in other countries include the Mary Kathleen mill in Australia [Baillie & Thomas 1972], Schwartzwalder mine in the USA [Seidel 1976], Besines mill in France [Michel 1976], Eldorado Beaverlodge and the Bicroft mine in Canada [Gow & Ritcey 1969].

Recent papers by Skorovarov *et al*. [1976] and Skrinichenko *et al*. [1977] described Russian practice and equipment. Ore is sorted with a practical upper limit on size of -300 +150 mm and a lower limit of -25 +5 mm. The ore sorted is maintained within a narrow size range with a ratio of upper/lower sizes of 1.5 to 2.0, which is achieved by screening. Production rates of separators currently in use are as follows:

	Separator Classification	Production Rate ($t\ h^{-1}$)
A	-200 +50 mm ore	40-100
B	-50 +25 mm ore	10-15
C	fine class	5

Up to 35 per cent of feed is rejected as waste rock. Two types of separators are described: one uses a cone distributing device to arrange the particles in single file for discrimination on an inclined chute, the other uses a vibratory tray distributor to set the particles in line on a stabilising belt, with the particles sorted in free fall and rejected by compressed air blast or water spray. In one example with ore containing 0.09% U, sorting in the various classes of machine gave a concentrate with an average grade of 0.13% U, with 95.4 per cent recovery of uranium and 33.7 per cent of the feed rejected.

Ore Sorters Ltd manufacture Mk VIA radiometric sorter machines for use with 50 to 150 mm and 25 to 50 mm feed sizes [Adorjan 1976]. These machines use compressed air to deflect radioactive minerals travelling on a belt moving at 1 $m\ s^{-1}$. The particles line up on the sorting belt in six parallel lines, with discrimination by a scintillation detector underneath each line. It is claimed that particles containing 0.1%U can be sorted with count times of the order of 0.1 s.

Grimes [1973] described radiometric sorting tests carried out on Rio Algom ores from the Elliott Lake area of Canada. Size limits on ore sorted ranged from 64 to 76 mm to about 150 mm. The results of tests by Ore Sorters Canada Ltd in 1969 on +76 mm Quirke mine ore showed that radiometric sorting was superior to heavy media and flotation methods for low percentage rejection, but was inferior at higher rejection. Calculated operating costs for a flowsheet incorporating radiometric sorting were higher than for the standard flowsheet.

4.1.2 Photometric sorting

In the early 1970s, the UKAEA developed an optical system, based on fibre optics, which differentiated between opaque, uranium-bearing particles and translucent grains of quartz or quartzite, and was claimed to be useful for eliminating them [Smith & Garrett 1972].

Scanning laser light sorters have been developed by Ore Sorters Ltd [Adorjan 1975, 1976]. A model 13 unit has been in operation for over four years at Doornfontein Gold mines, South Africa. The more advanced model 16, which uses multi-level reflective analysis, is capable of sorting 60 t h^{-1} of 20 mm feed.

4.1.3 Conductimetric sorting

Gunson-Sortex (Mineral & Automation) Ltd developed a sorter for the 50 to 150 mm size range which sorts on the basis of the electrical conductivity of the particles [Adorjan 1976]. This type of sorter has been used previously with Rio Algom Nordic ore [Grimes 1973]. In this application the selectivity was similar to radiometric sorting but because the ore had to be washed and dried it was ruled out on the basis of operating costs.

4.2 Flotation

For some South African Witwatersrand ores, flotation can be used to produce a concentrate containing up to 80 per cent of the gold and pyrite and 25 per cent of the uranium in less than 10 per cent of the original mass. This concentrate is finely milled and high recoveries of uranium and gold are obtained. The bulk of the ore is not milled and cost savings result [James 1976].

Three flotation plants with a combined treatment capacity of 50 000 t d^{-1} will be used to treat residual slimes and current residues for the recovery of sulphuric acid, uranium and gold, at Anglo American's President Brand plant in South Africa [James 1976; Payne 1977; Ruhmer *et al*. 1977]. The pyrite concentrates will be fed to a leaching plant for the recovery of uranium and then to the acid plant. The completed plant will come into operation in 1978 [WM 1977b]. The project is economically justified by the value of pyrite, gold and uranium recovered and by removal of old dumps from areas which have become valuable as real estate.

Flotation testing of Canadian ores has not produced satisfactory results [Grimes 1973]. For example, the flotation of uranium from Quirke mine ore using fatty acid reagents was shown to be relatively slow and, compared with radiometric sorting and heavy media separation, more selective, but also the most expensive.

A significant amount of unreported flotation work has also apparently been conducted on high-lime uranium ores in the US, one objective of which was the production of high and low lime fractions that could be treated separately [Seidel 1976]. Although previous results were not particularly favourable there is new interest in this approach as a result of the improved uranium price.

4.3 Heavy Media Separation

Heavy media plants have operated previously in South Africa at Anglo American Corporation's Free State Saaiplaas and Vaal Reef mines [James 1976]. Difficulties with abrasion and medium loss were encountered at the former plant. The Vaal Reef plant ran for about two years but was uneconomic.

The proposed flowsheet for the large scale processing of low-grade bituminous shales at Ranstad in Sweden includes a heavy media separation plant in which limestone inclusions are separated from the shale [Andersson 1976].

4.4 Gravity Concentration

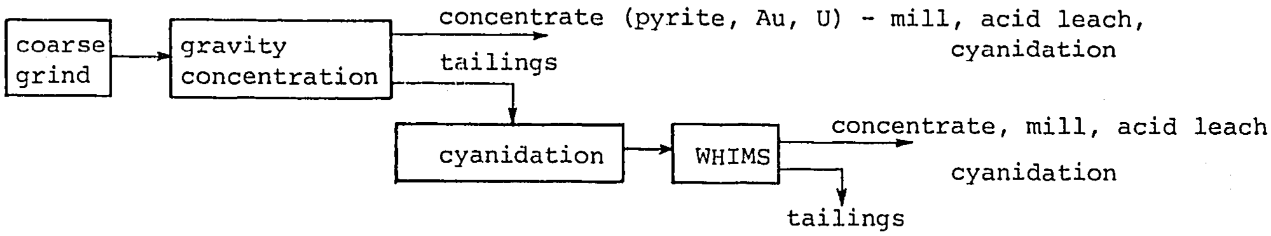
The Cluff Lake mine in Canada, due to start production in 1979 but now delayed pending an environmental inquiry, will have a gravity concentration plant at the mine [Lendrum 1976]. The uranium content is an average of about 20 per cent and the ore is a magma of massive uraninite allied with a matrix of chlorites. The gravity concentrate should contain up to 56% U.

Gravity concentration is also gaining popularity in the mineral industry where it is used to recover valuable heavy minerals from flotation plant tailings. In South Africa, the plant of the Palabora Mining Co. Ltd has used Reichert cones to recover a uranothorianite concentrate from copper concentrator tailings since mid-1971 [Nel 1972].

4.5 Wet High Intensity Magnetic Separation (WHIMS)

Some of the uranium-bearing components of the Witwatersrand ores are feebly magnetic and WHIMS is being developed as a method for recovery of uranium from low grade ores and gold plant tailings previously considered uneconomic [NIM 1975; James 1976]. Laboratory results on a large number of low-grade Witwatersrand ores and gold plant tailings have shown that up to 70 per cent of the uranium and 50 per cent of the gold can be recovered in a magnetic concentrate which in many instances represents less than 10 per cent of the feed. Considerable developmental work is said to be necessary to reproduce the results industrially but the method appears promising.

Research has also been carried out in South Africa on a new process for the treatment of Witwatersrand ores which includes gravity concentration and WHIMS steps [NIM 1975; James 1976] as outlined below:



This process offers, among other advantages, recovery of a considerable proportion of the uranium in ores that are too low in grade to be treated directly by acid leaching. For example, treatment of a feed containing 0.011% U gave 64 per cent recovery of uranium in the gravity and magnetic concentrates.

Uranium minerals in some Canadian uranium ores are also reported to have concentrated in the magnetic fraction when a dilute ore slurry was passed through a Jones wet magnetic mineral separator [Grimes 1973]. Tests on Rio Algom's Quirke mine ore using WHIMS showed that both finer grinding and multiple passes of reject were required to obtain acceptable rejection, and then only at high uranium loss. This loss plus the high capital cost of multiple units needed to treat the reject ruled out use of the method.

5. LEACHING PROCESSES

Interest in processing lower grade and more refractory uranium ores has led to investigations of alternative leaching techniques, and numerous review articles summarising recent research and development have been published during the last five years [Boutonnet 1975; Garrett 1975; James 1976; Leger & Boutonnet 1973; Lendrum 1974; Seidel 1976; Smith & Garrett 1972; Skorovarov *et al*. 1976]. The techniques receiving increased attention include strong acid cure leaching, ferric sulphate leaching, pressure leaching and ammonium bisulphate leaching.

5.1 Strong Acid Leaching

Strong acid leaching (SAL) involves impregnating coarsely ground, dry ores with a small volume, typically around 10 vol.%, of concentrated sulphuric acid (only sufficient to wet the ore) and curing at 65-100°C for 12-24 h. The uranium solubilised during the curing stage can be extracted from the ore by any of the following methods [Skelton & Scowen 1977]: deep bed percolation through the cured ore in a storage silo, washing the ore as a shallow bed on a belt or pan filter, or reslurrying and solid-liquid separation by conventional methods.

The leaching conditions of the SAL process are much more severe than dilute acid, slurry leaching. They are attractive for refractory ores which require a high residual acid concentration (up to 100 g ℓ^{-1}) to achieve adequate uranium extraction because the excess acid, present in the slurry, is considerably reduced. In addition, the hot concentrated acid is a sufficiently strong oxidising agent to eliminate the need for the addition of an oxidant. No fine grinding is required as SAL extracts uranium from ore at much larger particle sizes, which also simplifies the subsequent solid-liquid separation step. The treatment of a thick slurry by continuous agitation in pachucas can be replaced by a static curing process which may frequently require a shorter residence time. The quantity of water required for washing the coarse ore is less than for a fine pulp, and a stronger pregnant solution can be produced which allows a reduction in size of extraction equipment. The smaller volume of water required also makes it particularly suitable for applications in arid regions. The costs of neutralising the barren liquor are reduced because of the reduced acid consumption [Skelton & Scowen 1977].

The biggest disadvantage of SAL is that it requires almost dry ore for the leaching step which can be achieved only by dry grinding or drying after wet-milling.

A number of processes incorporating SAL have been reported. The UKAEA has patented a process [e.g. Smith *et al.* 1972] which has been the subject of extensive tests [Skelton & Scowen 1977]. Laboratory tests using 3M acid with conglomerate Canadian ores and sandstone ores from the Western United States, ground to about -1 mm, yielded extractions of uranium comparable to those obtained in dilute slurry leaching with similar total acid consumptions in a significantly shortened reaction time. An economic evaluation of the UKAEA process predicted that capital and operating costs would be in the order of 25 per cent lower than those for a conventional plant. The applicability of the UKAEA process to Australian ores with a matrix of a quartz-chlorite schist was also examined, but SAL was not suitable for this type of ore [Alfredson *et al.* 1975; Skelton & Scowen 1977].

Pechiney Ugine Kuhlmann also patented a similar process and were responsible for the design and construction in 1970 of a milling plant using this technique at Arlit in the Niger Republic [Boutonnet 1975; Leger & Boutonnet 1973]. The Arlit ore contains organic compounds of tetravalent uranium which require a free acid concentration of 100 g ℓ^{-1}

for satisfactory leaching. The acid consumption under conventional leaching conditions was 150 kg t^{-1} ore with a water requirement of 1000 ℓ t^{-1} ore. SAL gave an acid consumption of 55 kg t^{-1} ore when water was restricted to 50 ℓ t^{-1} ore. In addition, no oxidant was required and no external heating was necessary.

The Arlit process involves dry semi-autogenous grinding, with air at 70-80°C being blown into the ore during grinding. Unlike the UKAEA process which claims to be effective for some ores with a large particle size (-2 +1 mm), the Arlit ore requires grinding to < 2 per cent plus 800 μm to achieve satisfactory extraction. After grinding, the acid impregnation and curing steps are carried out in a rotating horizontal drum, the ore is then disintegrated in a revolving tube loaded with pebbles, and transferred to stirred tanks for repulping. The resultant slurry is separated by a rake classifier into sands and fines (-150 μm), which are washed separately in classifiers and thickeners, respectively.

Pechiney Ugine Kuhlmann have also proposed a similar process for a mill at Akouta in Niger which is scheduled to start up this year. Unlike the Arlit mill, curing will be effected on a patented, continuous rubber conveyor belt, following practice in the fertiliser industry, and washing and liquid-solid separation will be carried out on a vacuum belt-filter without previous separation of sands and fines [Crawford 1977].

5.2 Ferric Sulphate Leaching

Acidic ferric sulphate leaching is carried out at a higher temperature and with a higher concentration of ferric ion than are normally used in conventional dilute acid leaching. The spent ferrous sulphate solution is re-oxidised in a separate step and the reagent, after the addition of a small quantity of acid makeup, is recycled. Two independent processes incorporating this basic technique have been developed.

After comprehensive investigations of the reaction mechanisms and kinetics of the leaching of uraninite [Laxen 1973; Needes & Nichol 1972; Nichol & Needes 1972; Nicol *et al.* 1975], the costs were assessed of increasing temperature and ferric ion concentration (Felix process) to improve the extraction of uranium from South African ores [James 1976]. Tests showed that an iron concentration of 7 g ℓ^{-1} achieved an extraction of uranium of 93 per cent compared with 80-85 per cent obtained by conventional leaching using manganese dioxide as oxidant. It was originally envisaged that ferric ion would be recycled *via* autoxidation of barren solution with sulphur dioxide, but recently a technique has been developed for the oxidation of acidified ferrous sulphate solution by a bacterial film in the presence of dissolved oxygen; this technique eliminates the need for sulphur dioxide and is very cheap. James [1976] expected that at least one South African mill would soon convert to a Felix leach, with a wider application to follow.

A South African company, General Mining, has announced plans to install a unit (named Bacfox) at its Buffelsfontein gold/uranium plant to oxidise ferrous iron by bacteria. Pilot plant results indicated that the new process should increase extraction of uranium from 76 to 80 per cent and also cut operating costs by 8 per cent. In the Bacfox system, half of the filtered leach liquor passes through a tank where ferrous is oxidised to ferric ion and the liquor recycled to the leaching stage. In the tank, bacteria are grown on a honeycomb-like block of ribbed plastic which is completely submerged in the ferrous sulphate solution. Atmospheric air is sparged into the solution by an aerator which also ensures agitation. The temperature in the tank is maintained at 30°C and the pH at 1.8 [SAMEJ 1977a].

The UKAEA and the Warren Spring Laboratory have also demonstrated a ferric sulphate leaching process which arose from a study to develop bacterial leaching as an economically attractive in-plant process suitable for treating ores normally extracted by conventional dilute slurry leaching. It was originally planned that leaching of uranium would be carried out by reagents generated *in situ* by bacterial action on pyrite in the ore, but the concept evolved to include the production of leach liquor, in an external vessel, by the bacterial oxidation of a slurry of pyrite. Although useful leach solutions were produced, leaching rates were too slow. However, significantly greater rates were obtained with acid ferric sulphate at 50°C, pH 1.2, a ferric ion concentration of 12 g ℓ^{-1}, and a particle size of less than 3 mm [Derry *et al*. 1977].

A process flowsheet was developed and tested in pilot plant trials. Leaching of uranium was by an upward flow of acid ferric sulphate solution through fully submerged beds of coarsely ground ore contained in vats. The required residence time for Elliot Lake ore was ten days. The leachant was regenerated by oxidation of ferrous ion in a bubble column with the aid of bacteria which formed a film on the column walls. Retention time was about 12 h at 30°C and pH 1.7 [Derry *et al*. 1977].

The process is considered to have several potential cost advantages over dilute slurry leaching methods. The use of a bacterially generated leach solution yields higher extraction rates with a smaller acid requirement compared with those of conventional leaching. Fine grinding is avoided because a coarse particle size can be treated and liquid-solid separation is simplified. Simple equipment is used for the leaching and bacterial regeneration processes.

5.3 Pressure Leaching

In the USSR, an autoclave system is being successfully used to process refractory and complex low-grade uranium ores. The higher temperatures and pressures, together with oxygen as oxidant, achieve improved extraction rates of uranium for lower consumptions of sulphuric acid [Skorovarov *et al.* 1976].

Two types of autoclaves are used; the four-chamber horizontal type with a mechanical mixer and, more recently, a vertical type with pneumatic pulp stirring. Typical horizontal autoclaves have a capacity of 125 m^3 and are equipped with two-stage turbine mixers which individually suck up and disperse about 100 m^3 of air per m^3 of pulp. However, this type of mixer has proved to be very costly and accounts for about 40 per cent of the cost of an autoclave. The absence of rotating mixing devices simplifies the construction of the vertical type of autoclave, resulting in a decrease in construction and maintenance costs, and energy consumption. Typical industrial units have a capacity of 100 m^3 and are fitted with an airlift column along the central axis of the vessel which is used for pulp circulation. Air is supplied through a perforated plate, fitted with hoods, which is situated below the air lift column.

A process for the pressure leaching of South African ores at elevated temperatures, with air as oxidant, which resulted in the decomposition of pyrite in the ore to ferric sulphate and sulphuric acid, was successfully demonstrated in the 1950s. This method of treatment was not adopted because of the high capital cost and problems envisaged in transferring abrasive and corrosive pulps at high temperature and pressures. Recently, pressure leaching was reassessed and a demonstration plant was constructed comprising a 20 m^3 capacity, four-compartment, mechanically agitated autoclave and a spiral heat exchanger [James 1976].

A new pressure leaching process for treating refractory uranium ores has been developed recently by Lurgi Chemie and Huttentecknik in cooperation with Vereinigte-Aluminium-Werkel (VAW). The system employs a plug-flow 'tube digester' with a double tube. The slurry is introduced into the central inner tube whose first section is preheated by a countercurrent flow of hot slurry flowing in the surrounding outer tube while its last section is heated further by a heat exchanger using steam or fused salts. Advantages claimed for this system over conventional autoclaves are improved heat transfer coefficients, virtually no moving parts, faster reaction times as higher temperatures can be used, and rapid startup and shutdown. The construction of a pilot plant to treat a variety of ores has commenced and it is expected to commence operation shortly [NF 1977a].

5.4 Ammonium Bisulphate Leaching

The ammonium bisulphate leaching process was originally developed on a small scale by Pechiney Ugine Kuhlmann for the treatment of uranium bearing phosphates in the Central African Republic. An industrial pilot plant is now being developed [Boutonnet 1975]. In this process, ammonium bisulphate is used to digest the ore and, after solid-liquid separation, ammonia, followed by ammonium carbonate, is used to neutralise the solution and form ammonium sulphate. The ammonium sulphate is then decomposed by heating to regenerate the bisulphate for leaching and ammonia for neutralisation. Uranium is recovered by either solvent extraction before neutralisation, or thermal decomposition of the ammonium uranyl-carbonate precipitate.

5.5 Sodium Carbonate/Bicarbonate Leaching

Marginal improvements are being made in optimising alkaline leaching conditions and equipment at mills in Canada and USA [IAEA 1976a].

5.6 Control Instrumentation for Leaching

If the full potential of improved (or even conventional) leaching processes is to be realised, the concentration of acid, ferric and ferrous ions in the leaching slurry must be accurately controlled. Plant practice in South Africa showed that conventional equipment for measuring pH and redox potential was generally unreliable and improved methods were examined [Sommer *et al.* 1973; James 1976]. An electrodeless conductivity meter with a probe consisting of two toroidal coils spaced on a non-conducting tube, was developed to control sulphuric acid additions. The instrument was calibrated *in situ* by direct comparison with routine acid titrations [Sommer *et al.* 1973]. At one plant, where the instrument was subjected to full-scale trials, the free acid concentration was kept between 2 and 10 g ℓ^{-1} by normal methods of measurement and control. The new automatic control system using the conductivity meter succeeded in maintaining the free acid level at 6.8 ± 0.4 g ℓ^{-1} with considerably reduced sulphuric acid consumption [James 1976]. More recently, the Highland uranium mill in Wyoming has also installed a conductivity-sensing unit for concentrations of 0-5 per cent sulphuric acid. The unit is calibrated by using pH meters for grab-sample reference control [ME 1978].

Another development was a more robust and reliable electrode assembly for redox measurements. It consisted of a Lazaran reference electrode, a silver-silver chloride type totally encased in polymer and a plastic encased platinum electrode [Sommer *et al.* 1973].

6. *IN SITU* LEACHING

The substantial increase in the price of uranium over the past few years has stimulated interest in the recovery of uranium from small and low grade ore deposits by *in situ* leaching techniques. These include the use of solutions to extract metal values from broken or unbroken rock within the orebody (solution mining), from broken rock in underground mines (stope leaching), and from broken rock within open cut mines or after removal from underground mines (heap leaching). The last two techniques are frequently enhanced by bacterial action. Much of the information on these processes is reported at the annual meetings of the American Institute of Mining, Metallurgical and Petroleum Engineers [e.g. AIME 1974, 1976, 1977]. Bibliographies on *in situ* leaching of uranium (and copper) have been compiled by USAEC [1974b] and Larson & Dandrea [1976].

6.1 Solution Mining of Permeable Ores

6.1.1 General principles

Solution mining has been adopted to exploit copper, evaporites, gold and uranium deposits in the United States, Canada [Bhappu 1976] and USSR [Skorovarov *et al.* 1976]. For those orebodies which are amenable to this technique, it has the advantages, over conventional mining and milling, of

- low initial capital investment,
- short lead times,
- low manpower requirements,

- less environmental impact (in terms of surface disturbance, tailings storage and radon release) and,
- compared with underground mining, the whole of the orebody can be utilised [Bhappu 1976; Hancock 1977; Hunkin 1975; Lackey 1975].

The disadvantages are

- difficulty in controlling and predicting groundwater hydrology,
- ensuring containment of leach solutions in the orebody,
- restoring the original groundwater conditions after cessation of mining and,
- monitoring the whole operation [Rouse 1974; Shock & Conley 1974].

Where the orebody is permeable, e.g. some sedimentary deposits, the leaching solution may be injected into the ore through boreholes or wells and withdrawn at other wells to control the underground flow of leach solution and the extent of metal recovery. The following criteria generally apply [Anderson & Ritchie 1968; Lackey 1975; Hancock 1977]:

(i) The orebody should be a generally horizontal bed underlain by a relatively impermeable stratum.
(ii) The orebody must be below the static water table.
(iii) The direction and velocity of water flow must be known.
(iv) The ore must be amenable to the proposed leaching agent.

Obviously, well spacing is a vital factor [Seidel 1976] and many patents and techniques have been reported for processes which facilitate this type of operation by controlling and confining the underground flow of leach solution to minimise loss of leach solution and contamination of groundwater [Rhoades 1973, 1974]. Mathematical modelling has been used to investigate various geometric patterns for injection and production wells [Hancock 1977; Shock & Conley 1974]. A uniform square array of wells, with an injection well at each corner and a production well in the centre, is generally used [Hancock 1977]. For example, the US Steel-Niagara Mohawk operation at Clay West, Texas, has a square pattern of injection wells on 15 m centres and production wells at the centre of each square [Davis *et al.* 1976; White 1975]. However, the Intercontinental Energy Company uses a computer design procedure which takes hydrogeological conditions into account and generates a non-uniform pattern of wells for maximum sweep efficiency and control of leachate [EMJ 1977a].

The leaching operation is controlled by balancing injection and production rates. Monitoring wells around the leaching area are sampled regularly to detect any migration of injected solutions which, if detected, is counteracted by adjusting the injected and produced volumes. Upward migration is minimised by casing and cementing all wells, and shallow monitor wells are used to detect such leakage.

6.1.2 Choice of leachant

For solution mining, the choice of leaching agent depends on the mineralogy of the ore and gangue and its selectivity for uranium. Residual reagents and soluble products (including losses), after exploitation is completed, should not affect groundwater quality beyond permissible limits. The effect of the reagent on the permeability of the underground formation is more important than its uranium extraction efficiency. Both acid and alkaline leaching flowsheets have been used but the latter are more common [Hunkin 1976; Hancock 1977]. An oxidant is also required e.g. oxygen, sodium chlorate or hydrogen peroxide.

Sulphuric and nitric acid leaching have been used in the United States [Anderson & Ritchie 1968] and sulphuric acid has been used in Czechoslovakia [Hunkin 1976]. Sulphuric acid tends to reduce permeability by the formation of silica gel and precipitation of calcium sulphate from clays and calcite, and it dissolves other metallic minerals, e.g. V, Mo, As, Fe which are associated with uranium. Because of this, sulphuric acid requires more surface processing and greater controls than alkaline solutions to maintain the water quality of aquifers [Hunkin 1976].

Carbonate leaching, under conditions of moderate pressure (0.5-1.5 MPa), temperatures in the range 10-75°C and low concentrations, does not suffer from these disadvantages. In particular, the ammonium ion is preferred (e.g. at the Wyoming Mineral Corporation's Irigaray project), because it produces less swelling of some classes of clays, but it may be difficult to reduce the ammonia concentration to acceptably low levels after operations cease [NF 1978a]. The use of sodium bicarbonate may be more advantageous in the future, particularly because of its lower cost [Hunkin 1976].

6.1.3 Uranium recovery and waste disposal

The uranium in the leach solution, typically 10 to 100 mg ℓ^{-1}, may be recovered in various ways, depending on the scale of operation, proximity to a conventional mill and uranium concentration. For a large operation, independent processing to yellowcake may be desirable. For small orebodies adjacent to an existing mill, the solution may be concentrated by ion exchange and combined with the main mill streams.

The leach liquor is not turbid but may require simple filtration through a sand filter or carbon columns before ion exchange. Ion exchange systems employing fixed beds or continuous contactor designs, as used by the US Bureau of Mines, Higgins, Cloete-Streat, and Himsley and Porter have been used or are under consideration (see Section 3.6.3) [Hunkin 1976]. Chemical wastes from the ion exchange recovery plant are concentrated and retained in evaporation reservoirs lined with a chlorinated polyethylene lining at the Clay West operation [White 1975]. Present plans for final restoration are to pump solutions out of the orebody into deep disposal wells until ion concentrations are reduced to acceptable levels.

6.1.4 Applications

Solution mining is still very much at the developmental stage and considerable additional research is required to optimise the technology and provide confidence in a proposed operation before it commences [Frank 1975]. Bhappu [1976] outlined a systematic approach for feasibility studies of proposed ventures.

NEI [1977] reported that 15 or more *in situ* leaching test programs are in progress in USA. The major operations in permeable orebodies are listed in Table 2, and have design production rates of over 85 tU/year. The Clay-West mine of US Steel-Niagara Mohawk has been described in detail [Davis *et al*. 1976; Narayan & Rand 1976; White 1975] and Bhappu [1976] has listed other solution mining projects.

6.1.5 Economics

Hunkin [1975] reported solution mining costs of \$23.1/kg U compared with \$29.6/kg U for a conventional mine and mill and also stressed the lower initial capital investment (approximately one fifth) and the much shorter lead time (approximately one third) for solution mining.

Frank [1976] described a cost model for solution mining of uranium with annual production rates of 100 to 1000 t U, corresponding to leach solution flowrates of 125 to 500 $\ell\ s^{-1}$. A detailed cost breakdown was given for the particular case of 290 t U/year with capital and operating costs of \$14.3 and \$23.9/kg U, i.e. a total of \$38.2/kg U. Corresponding total costs for production rates of 190 and 380 t U/year were respectively \$44.7 and \$36.1/kg U.

The most important cost items were mine operations (injecting leach solutions to the mineralised zone and recovering the leach liquor), mill operations (processing leach liquor, analysis of process solutions and drilling of waste disposal wells), and primary development (drilling of wells, casing, cementing, etc. and drilling of monitor holes). Automation of the mine operations, thereby reducing manpower requirements and more efficiently using power and chemicals, and optimising the drilling, development and spacing of wells may lead to a substantial reduction in the costs of solution mining.

Males [1977] summarised research on cost models being developed for the Electric Power Research Institute. As an example of typical costs for the case of a 6 m thick ore deposit (0.05 per cent grade) at a depth of 180 m giving 70 per cent uranium extraction and 15 per cent return on investment, he quoted a cost of $68/kg U (in 1977 dollars). This is higher than the earlier estimates mentioned previously. More recently, WM [1978a] have published a set of twelve graphs which provide a consistent framework for determining the relative cost ranges for uranium solution mining projects. The variables taken into account include uranium production rate, leaching solution flowrate, uranium concentration in pregnant liquor, injection rate per well, number of well field patterns, depth of wells, life of operation, initial exploration, development and drilling costs, efficiency of uranium extraction, ore grade and royalties.

6.2 Solution Mining of Broken Orebodies

If natural permeability does not allow satisfactory percolation rates through the orebody, fracturing may be necessary using hydraulic techniques or chemical (possibly nuclear) blasting. Hydraulic fracturing, by pumping high pressure water through a formation, is a well established and extensively used means of stimulating oil and gas wells and is equally applicable to solution mining operations. Porter & Carlevato [1974] described chemical blasting techniques to prepare orebodies for *in situ* leaching, and McLamore [1974] outlined the potential use of liquid explosives which can be pumped.

With deeply buried deposits, it is proposed to break the ore and produce chimneys of highly permeable rubble surrounded by fracture zones. Distribution and collection galleries are then used to distribute the flow of leach solution. A great deal of information has been published on the possible use of nuclear explosives for breaking such ores, e.g. Lewis *et al.* [1974], Bhappu [1976], but this method has not been used commercially.

In shallower or surface deposits, the broken rock may be exposed at the surface. Acid leach solution is then sprayed on the surface and seeps through the ore to the collection gallery. These techniques have not generally been applied to uranium ores but various applications to copper extraction were described by Bhappu [1976] and Lackey [1975].

6.3 Stope Leaching

Significant quantities of low grade ore which remains in stopes in a mine because it cannot be recovered economically may also be recovered by *in situ* leaching techniques. Subsidence or caving in of underground mines is generally not intentional but can also enable the resulting broken rock to be leached by similar techniques. The acid leach solutions can generally be arranged to drain into other stopes and shafts leading to lower levels in the mine and then pumped to the surface for recovery of the uranium. This type of operation represents a low capital cost method of treating residual, low grade, broken ore and is used in Canada for uranium recovery [DEMRC 1977; IAEA 1976a; Lendrum & McCreedy 1976].

Since the size range of broken ore is similar to that from conventional mining, the technology developed for heap leaching (Section 6.4) is, essentially, applicable in stope leaching. Avoiding chemical precipitation is not as critical with broken rock as for unfractured orebodies and consequently sulphide ores can be treated by this method. In addition, bacterial leaching will enhance the leaching of sulphide ores. As for solution mining, stope leaching also relies on the underlying rock strata to contain the leaching solution and minimise losses.

Kerr Addison Mines commenced operation of a 500 t/year mill at the Agnew Lake uranium mine in 1977 which uses leaching of low grade (0.042% U) ore in stopes and in surface stockpiles. The test work leading to this exploitation included an underground test on a 45 t block of broken ore using leaching with a bacterially active sulphuric acid medium which produced 70 per cent uranium recovery in a little over six months [Lendrum & McCreedy 1976]. A Himsley continuous ion exchange contactor (see Section 8.3) is used for recovery of the uranium from the leach liquor. Capital cost was about 40 per cent of that for a conventional uranium mill [NF 1978b]. A recent report [MJ 1978] indicated that production in 1977 was well below design capacity which will not now be reached before 1979.

6.4 Heap Leaching

Heap leaching of uranium ores with acid is currently practised in the United States [NF 1977b; Facer 1977; Merritt 1971; Seidel 1976], France [Michel 1976, 1977), Spain [Gomez Jaen *et al*. 1977; Josa 1976; Merino & Josa 1977], Portugal and Argentina [OECD/IAEA 1973]. Pilot plant scale heap leaching tests have also been carried out in India [Jayaram *et al*. 1976] and Mexico [Ajuria-Garza *et al*. 1975; Ajuria-Garza & Jamrack 1976]. Heap leaching practice was reviewed by Merritt [1971] and more recently by Lowson [1975] and Michel [1976, 1977]. Typical applications include [Michel 1976]: heap leaching of marginal or low-grade ores as mined in minimum cost operations on areas sealed with plastic sheeting; of run-of-mine ores, after crushing, on recoverable areas such as bitumen pads; and accelerated leaching in vats of average grade crushed ores with limited reserves.

Woolery *et al*. [1977] compiled a history of heap leaching at Gas Hills, Wyoming, beginning with laboratory and pilot plant tests and leading to construction of a 204 000 t pilot heap in 1973-1974. On the basis of this experience, the Maybell heap leaching operation comprising a heap 200 m long x 150 m wide x 6.4 m high, and a skid-mounted processing plant was constructed in 1976-1977 to treat low grade ore averaging 0.025% U at an abandoned Union Carbide Corporation mine and mill site in Colorado.

In the US, at least two projects are under way to recover uranium from tailings piles at inactive mill sites by heap leaching [NF 1977b]. Solution Engineering and Texas utilities will leach tailings in place at the inactive Falls City, Texas, uranium mill. Ranchers Exploration and Development Co. is constructing a plant, near Naturita, Colorado, for recovery of uranium from 545 000 t of purchased tailings [EMJ 1976; MJ 1977b; NF 1977b] containing about 0.042% U which will be moved to the new site for acid leaching in impervious earthen tanks. Preliminary estimates suggest that uranium recovery would be about 50 per cent. A second operation is planned for Durango, Colorado, using 1 270 000 t of tailings containing about 0.034% U.

Leaching of crushed ores in vats allows treatment over shorter periods and usually gives higher recoveries. Vat leaching in rubber lined ditches containing about 2000 t of ore has been practised at Langogne, France. The time for charging, leaching and washing of a heap was about one month. The vats have recently been replaced by bitumen pads [Michel 1976, 1977]. The leach solutions are purified by solvent extraction or ion exchange and the uranium precipitated with alkali as in conventional mills, often in mobile, skid-mounted units.

6.5 Bacterial Leaching

There is also a continuing interest in the application of bacterial leaching to uranium ores, and research is continuing into the amenability of particular ores [Ebner 1972; IAEA 1972, 1974; Jayaram *et al.* 1976; Nerkar *et al.* 1974] and into the more basic mechanisms and kinetics of bacterial oxidation processes [Ebner 1972; IAEA 1974; Josa 1976; Khalid & Ralph 1977]. Lowson [1975] reviewed all aspects of bacterial leaching. At the present time, leaching of uranium is limited to pyritic ores where autotrophic bacteria oxidise the pyrite to sulphate and ferric ions. The orebody becomes progressively more acidic and oxidising until the uranium is solubilised. Most success has been obtained with the bacteria *thiobacillus thiooxidans* and *thiobacillus ferrooxidans* which grow in strongly acidic media. It is possible that, with development, the method might be extended to higher pH and even alkaline conditions. The IAEA has maintained a coordinated research program on bacterial leaching of uranium ores since 1972 [IAEA 1972, 1974; Lowson 1975].

The actual mechanism by which the bacteria achieve mineral oxidation is not fully understood and elucidation of the reactions may enable the tailoring of bacteria to particular environments.

The bacterial action can be improved with nutrients [Jayaram *et al.* 1976; Lowson 1975] but it is not always economic to do so. Recent work in India on quartz chlorite schists showed that 9K nutrients helped to reduce acid consumption and leaching time and increase uranium solubilisation [Jayaram *et al.* 1976]. Neither seeding the bacteria nor feeding nutrients continuously was necessary; use of the nutrients in the initial batch and the ion exchange barrens in subsequent cycles was sufficient. By comparison the use of nutrients in the Canadian bacterial leaching operations at the Milliken, Stanrock and Denison mines during the 1960s was found either to be uneconomic or not to increase uranium recovery appreciably [Lowson 1975].

Bacterial leaching has been practised on a large scale in Portugal, Canada and Hungary in connection with stope and heap leaching operations (see Sections 4.4.3, 4.4.4) [IAEA 1974; Lowson 1975].

In Australia, recent research has been directed to the problem of heavy metal pollution (mainly copper and zinc) resulting from bacterial oxidation of metal sulphides in the waste heaps from the previous uranium mining operations in the Rum Jungle area. Microbiological studies have been undertaken to contribute to the understanding of the basic physical, chemical and biological mechanisms [Khalid & Ralph 1977] and modelling of heap behaviour has been attempted [Ritchie 1977; Ritchie & Clancy 1975].

7. SOLID-LIQUID SEPARATION

The two conventional solid-liquid separation systems in uranium ore processing plants are countercurrent decantation in thickeners, and filtration. Also in use are combinations of thickeners with cyclones and classifiers, the slimes being treated in the thickeners and the sands in either the classifiers or cyclones. Sand-slime separation and washing using cyclones and classifiers followed by resin-in-pulp uranium recovery (Section 3.6) have also been practised. Final clarification before uranium recovery by ion exchange or solvent extraction is accomplished typically by sand, pre-coat or leaf filters [IAEA 1976a].

Solid-liquid separation equipment represents a significant part of the capital cost of a uranium mill and considerable effort has been devoted to improved design and performance.

7.1 Thickeners

Recent developments are the Enviro-Clear and lamella thickeners. Enviro-Clear thickeners were initially developed for waste water treatment and are being evaluated for solid-liquid separation in uranium circuits. The flocculated feed slurry is introduced into a previously formed sludge bed rather than above the bed as in a conventional thickener. A fivefold or greater reduction in required settling area has been reported for some materials [Enviro-Clear 1976; Seidel 1976]. The first plant application is at the Wyoming mill of Bear Creek Uranium where five Enviro-Clear thickeners have been installed [WM 1978b]. Their performance, to date, has been equivalent to that of larger conventional thickeners. Lamella thickeners incorporate a nest of inclined parallel plates, situated in a sedimentation chamber, which reduce settling distance and increase effective area [Adorjan 1976; Sala 1974]. They are claimed to require less than one fifth the space needed by conventional thickeners. A lamella thickener, 2.2 m wide x 5.2 m long x 4 m high, has replaced a 17 m diameter single-stage lime slurry thickener at Eldorado Nuclear's Beaverlodge uranium mill, with both process and economic benefits [Adorjan 1975; Roe 1977].

7.2 Horizontal Belt Filters

Horizontal belt vacuum filters are increasingly being employed in French designed uranium mills [Renaud & Boutonnet 1977] because they offer significant capital and operating cost savings. The first application to uranium milling was at the Forez mill in France where a belt filter was installed in 1973. The Akouta mill in Niger, which will start operation in 1978, will use belt filters for solid-liquid separation as will the Novazza mill in Italy, which is scheduled to start up in 1979. The decantation circuit at the Bessines mill in France was to be gradually replaced by belt filters from the latter part of 1977.

In addition to offering lower unit capital and operating costs, belt filters are claimed also to produce a more concentrated pregnant liquor than thickening, with resulting cost savings in the design of the solvent extraction section. Washing and power requirements are also significantly lower than for rotary drum filters. Belt filtration could potentially eliminate the need for conventional tailings ponds as the tailings produced are in a compact form and could be back-filled into tanks or heaped and continuously rehabilitated [Renaud & Boutonnet 1977].

The new Randfontein Cooke plant in South Africa, and the refurbished Millsite facility, commissioned in August 1977, both include horizontal belt filtration [SAMEJ 1977b]. These plants process gold-uranium ores and slimes from an old dam. The Cooke plant includes five 120 m^2 Delkor belt filters for gold and four for uranium, and the Millsite plant includes one for uranium.

8. URANIUM PURIFICATION

Uranium is usually extracted from the clarified liquor and partially purified by solvent extraction, resin ion exchange or a combination of both methods (resin exchange followed by solvent extraction). Solvent extraction is usually carried out in mixer-settlers and ion exchange in fixed or moving bed systems. The high capital cost of equipment required to produce a solids-free solution for solvent extraction and conventional resin ion exchange has resulted in the development of new or improved methods of accomplishing the purification step. In these methods, only the coarse sands are separated from the fine slimes. The pregnant liquor with a low solids content of slimes is then treated by a resin-in-pulp (RIP) or a solvent-in-pulp (SIP) system. Initially the RIP process was used with resin contained in baskets, and later as a continuous operation in which resin and pulp were passed countercurrently through agitated vessels [Merritt 1971]. In the first designs, the two phases were separated between each vessel by vibrating screens or settling cones but a number of multistage continuous contactors were developed in the 1970s.

8.1 Solvent Extraction Equipment

The mixer-settler equipment used in solvent extraction operations offers flexibility and few operational problems. Research and development during recent years have concentrated on modifying the basic design to reduce capital and operating costs [Browning 1972]. This work includes impeller design studies and the use of coalescers to improve phase disengagement [Seidel 1976] and modified designs to reduce the size of equipment. For example, a patent has recently been issued to Holmes and Narver Inc. for a baffled settler which could reduce the construction cost of solvent extraction plants by up to 25 per cent. The new settler tank has baffles that alter flow patterns, giving greater efficiency of separation, and thus reduce space requirements. Pilot plant tests have reportedly demonstrated that the size of a settler tank can be halved using this method [MJ 1977a].

In a conventional mixer-settler, the settler occupies a considerable surface area and must be installed, for hydraulic reasons, at the level of the outlet of the mixer tank. As the settler is usually less than 0.5 m deep, it must be raised up on an extensive and expensive support structure. The alternative of having the settler at ground level and the mixer sunk into a pit can produce cost savings, but at the expense of convenience in plant operation. In addition, the two requirements of the pumping mixer (pumping and mixing) incorporated in modern plants are not always compatible, as the energy input required to develop the necessary pumping head to overcome pressure losses in the system can result in the formation of a dispersion which is too fine for efficient separation in the settler. It is therefore highly desirable to separate the two functions as far as is practicable, and to reduce the pumping requirements of the mixer.

These drawbacks associated with conventional units are overcome by a recent design development known as the 'low-profile' mixer-settler.

In this the conventional mixing tank is replaced by a mixing chamber, where the required dispersion is generated, followed by a channel which has the same depth as the settler and which provides the residence time necessary for the chemical reaction to the place [Skelton & Scowen 1977]. The flow through the residence chamber is kept high enough to prevent premature disengagement of the dispersion; if necessary, low energy mixers can also be used. As the mixer has the same depth as the settler, the pumping head is substantially less than for a more conventional unit.

The low profile layout also allows the entire plant to be installed at ground level, which offers considerable savings in the cost of supporting structures and stiffening of settler bases without the inconvenience of housing the mixer tanks in a pit. A small trench is necessary to accommodate the inlet pipes to the mixer, but the amount of excavation work required is small. Gravity flow can still be maintained throughout the plant, though organic and aqueous return pumps are usually provided, and the need for underground tanks is avoided. The savings of total energy input are likely to be small, but the reductions in the pumping energy input will reduce the amount of secondary haze produced. A plant that utilises a low profile configuration has been in operation for several months and performance has exceeded expectations [Skelton & Scowen 1977].

A variety of columns and other solvent extraction equipment has been proposed for uranium extraction but the lack of scale-up data has limited consideration of these units for mills built recently [Seidel 1976].

8.2 Solvent-in-Pulp Equipment

Solvent-in-pulp processes have not been used on a plant scale, principally because emulsion formation often leads to excessive solvent losses through entrainment of small globules in the solids pulp. For most systems, the solvent loss (> 4.5 ℓ of solvent/t ore) is 4 to 8 times the losses in conventional circuits. Lucas & Ritcey [1975] reported that a number of surface-active agents are effective in reducing or inhibiting amine adsorption on slurry solids and the formation of emulsions. With this improvement, they concluded that solvent-in-pulp processes could now be economic.

8.3 Resin-in-Pulp Processes

Developments in resin ion exchange technology over the last five years have been mainly confined to contactors. Several continuous or semicontinuous multistage fluidised bed columns have been developed which show considerable promise in reducing the cost of uranium recovery. These are the US Bureau of Mines (USBM) column [Traut *et al*. 1978], the Cloete-Streat (developed by the National Institute for Metallurgy (NIM), South Africa [Haines *et al*. 1975] and the UKAEA [Streat & Qassim 1973]) and Himsley contactors developed in Canada by Himsley Engineering Ltd. [Seidel 1976; Lendrum 1974; Lendrum & McCreedy 1976]. The essential differences between these columns are the interstage plates and the manner in which resin is transferred between stages.

The USBM column uses a simple sieve-tray design which permits a reverse flow of resin when the feed stream is momentarily interrupted, while the UKAEA column uses a double sieve-tray design which requires a flow reversal to accomplish resin transfer. The NIM column also uses a sieve tray design but the bottom tray is fitted with caps, and the Himsley column uses an external piping system. All of these columns can handle slime slurries containing up to about 7 - 8 wt % solids. The upper limit is reached when the combination of slimes, slurry density, viscosity and velocity causes the resin to overflow with the slurry leaving the top of the column.

Recently, plans were announced to use continuous ion exchange (CIX) columns in conjunction with processing of gold-uranium ores in South Africa [EMJ 1977b; James 1976] and with recovery of uranium from low-grade solutions produced by *in situ* mining ventures in the United States [Seidel 1976; EMJ 1977b]. The CIX system would also be extremely suitable for the treatment of unclarified leach solutions generated by percolation washing of coarse ore which has been proposed for use in the strong acid leaching processes. Of these plans, those announced for South African mills are at the most advanced stage.

The development of the NIM/CIX column has been described by James [1976] and, following the successful operation of pilot and demonstration plants, the Blyvooruitzicht Gold Mining Company has announced plans to convert its entire mill to a countercurrent decantation (CCD)/CIX system [EMJ 1977b; SAMEJ 1977c]. Unlike other mills in South Africa which use rotary drum filters followed by a sand or pressure clarifier to produce a solids-free solution for subsequent solvent extraction or fixed bed ion exchange, the new system will use countercurrent decantation (CCD) and eliminate the expensive final clarification stage. The proposed CIX process will use a strong-base resin and elution with 10 per cent sulphuric acid followed by solvent extraction for final purification. Advantages claimed for the CIX process include a reduction in plant capital cost of 20 per cent compared with previous generation Purlex plants (clarification followed by solvent extraction) and a reduction in resin inventory by one half to two thirds of the amount required in plants using a fixed bed ion exchange system. Future improvements of the system are the use of weak-base resins (thus eliminating the SX stage), denser resins and pulps containing 20 wt % solids (rather than 5 wt % solids at present) and belt filters (in place of a CCD circuit).

A fluidised ion exchange process which operates in a system of multiple tanks has also been developed in South Africa. Resin and solution are contacted countercurrently in a series of individual absorption vessels and loaded resin is periodically removed from the first absorption tank by airlift. A five-stage version of this system capable of treating 45 m^3 min^{-1} was installed at the Rossing plant in South West Africa [Seidel 1976].

The use of RIP systems for treating dense pulps is well established in Russia where it has considerably improved the economics of processing low grade ores because it eliminates the filtration step [Laskorin *et al*. 1977]. Contacting equipment includes fluidised beds, multiple tanks with mechanical stirring, moving bed columns and vessels with a pneumatic system of mixing.

Although most effort has been directed towards the development of ion exchange equipment, there have been improvements in the types of resins available. Weak base resins (such as Rohm & Haas XE270 and XE299), which produce a high grade yellowcake without the need for SX, are in commercial use; however, these resins have a low specific gravity and cannot be used in fluidised bed columns. Various attempts have been made to produce a high density resin (1.25 - 1.3 g cm^{-3}), preferably the weak-base tertiary amine type [Streat & Callaghan 1973; Yarnell 1975] but, at present, none is available commercially.

9. YELLOWCAKE PRECIPITATION AND FINAL PROCESSING

The specifications for uranium concentrates (yellowcake) acceptable to refineries for further processing to uranium dioxide, uranium metal and uranium hexafluoride depend on the particular refinery flowsheet and its capability to remove certain impurities such as sodium, calcium, arsenic, boron, vanadium, sulphate and phosphate. Table 3 summarises the current specifications. In general, there has been a recent trend towards as high a grade product as possible, typically containing greater than 76 wt % U (90 wt % U_3O_8). This trend is the result of changes to the precipitation, dewatering, drying and calcination steps.

9.1 Precipitation

Ammonia is used widely in South Africa, Canada and USA for precipitation of uranium as a uranate (yellowcake) although, as in early uranium mills, magnesia and/or caustic precipitation is also used [Hartley 1972]. The modern use of ammonium sulphate in place of sodium chloride for stripping the loaded solvent in solvent extraction strongly favours the use of ammonia for yellowcake precipitation thus enabling a higher grade product to be obtained. This approach has been adopted for new mills such as Highland [USAEC 1973], Shirley Basin [USAEC 1974a], Bear Creek [USNRC 1977], Ranger [1974], Pancontinental [1977] and Queensland Mines [1977] mills.

Precipitation of yellowcake from less pure strip solutions, e.g. from sodium chloride ion exchange eluate, which may also contain significant concentrations of sulphate and vanadium, or from carbonate-bicarbonate leach liquors from alkaline leaching flowsheets, is carried out with caustic soda solutions. This product is less pure than that obtained by ammonia precipitation and, for example, its sodium and vanadium contents may exceed those allowed by the Allied Chemical Corp. specification. The product may be upgraded by dissolution in sulphuric acid and re-precipitation with ammonia [Hardy 1972; Merritt 1971]. Alternatively the use of hydrogen peroxide (to precipitate hydrated uranium peroxide) in lieu of caustic soda has been implemented by the American Metal Climax Co., Colorado, for a sodium carbonate strip solution from solvent extraction [Merritt 1971] as well as at the Atlas Minerals mill, Utah, [Caropreso & Badger 1973] for a sodium chloride ion exchange eluate. A higher grade product which is readily filtered and more dense (1.25 times) than that obtained by ammonia precipitation was claimed.

The Atlas process incorporated ambient temperature, batch precipitation (using 150 per cent of the stoichiometric H_2O_2 requirement) and the precipitate was filtered, washed and dried at 350 - 520°C to give a product from 76 - 81 wt % U. By comparison with ammonia precipitation the cost of chemicals for peroxide precipitation was slightly higher (7.8 ¢/kg U) but the concentrations of impurities were substantially lower and within the Allied Chemical Corp. specification.

Shabbir & Tame [1974] showed that peroxide precipitation was also applicable to a wide variety of eluate and strip solutions. Their recommended optimum conditions were generally similar to those used by Atlas. The excess peroxide required is believed to be decomposed by impurities or consumed in the formation of soluble complexes with vanadium, molybdenum, etc. Interference of vanadium in uranium precipitation is reduced by increasing the amount of hydrogen peroxide added.

9.2 Dewatering

Dewatering operations most commonly involve thickening to increase the concentration in the precipitated yellowcake slurry to 15 - 30 wt % before solid-liquid separation using either rotary drum vacuum filters or solid bowl centrifuges. The latter have been increasingly favoured in new plants [USAEC 1973, 1974a; USNRC 1976, 1977; Pancontinental 1977; Queensland Mines 1977; Ranger 1974]. Generally, solid bowl centrifuges enable a clearer filtrate and a lower moisture content in the dewatered cake, and enclosed operation minimises the possible spread of contamination [Janov *et al.* 1975].

9.3 Drying and Calcining

A variety of equipment has been used to dry the yellowcake precipitate and partially decompose it to an oxide product by calcination at temperatures up to 800°C. With ADU, most of the water, ammonia and nitrogen is driven off below 450°C, the UO_3 initially formed begins to decompose at 400°C and complete conversion to U_3O_8 occurs between 650 and 800°C. Complete dehydration of uranium peroxide hydrates occurs below 500°C and heating to higher temperatures forms U_3O_8.

While separate equipment can be used for drying and calcination, multiple hearth refractory-lined roasters are commonly used for both steps. Wyssmont Co. Inc., USA, developed an AISI 304 stainless steel 'Turbo-Heat-Treater' unit which was claimed to have the advantage of a very close temperature profile and no refractory requirements [EMJ 1975; Lee 1976]. The unit, which has a capacity of 10 t d^{-1} comprises a slowly rotating (< 1 rev min^{-1}) vertical stack of 10 to 15 trays housed in a insulated shell 3 m high x 2.9 m dia. The yellowcake is fed to the top tray, then transferred continuously to each succeeding lower tray in a plug-flow fashion with the help of stationary wiper blades and leveller arms. Intimate gas-solids contact is facilitated by a vertical cross-flow of hot air (from an external direct oil-fired heater) through the housing, coupled with fans rotating inside the tray stack. The unit is more compact than conventional roasters and avoids the problems of spalling and cracking of refractories when handling wet feed materials.

Drying and calcination operations are carried out under negative pressure with the exhaust gases passing through dust collection systems to minimise yellowcake losses. Typical dust collection systems comprise initial cyclones, filters and wet scrubbers to remove very fine particles. The 'Turbo-Heat-Treater' unit is claimed to minimise dust carryover by the inherently low velocity and low turbulence in the tray stack, a specially designed air exhaust system and an 'over the rim' design which precludes solids entrainment in the exhaust gases.

10. RECOVERY OF URANIUM AS BY-PRODUCT

The only well established large-scale production of by-product uranium has been from the treatment of gold ores in South Africa where the revenue from gold is approximately twenty times that from uranium [James 1976]. The projected increase in the demand and price of uranium has stimulated interest in by-product sources of uranium, particularly phosphate rock and porphyry copper dumps.

10.1 Uranium from Phosphates

The concentration of uranium in phosphate rock which varies from 0.003 to 0.02 per cent is one to two orders of magnitude lower than in usual uranium ores. However, as world production of phosphate rock is approximately 120 million tonnes per year, recovery of by-product uranium could provide a significant source of uranium if the phosphate values in the rock were recovered. Consequently, methods have been devised to obtain uranium as a by-product from existing processes for the production of fertilisers (e.g. single superphosphate, triple superphosphate, ammonium phosphates, etc.). With present technology, uranium can be recovered at a reasonable cost only from the manufacture of wet-process phosphoric acid which is required for the production of high analysis fertilisers such as triple superphosphate and ammonium phosphate. Over 50 per cent of the total world production of phosphate rock is used for phosphoric acid manufacture [Ring 1975, 1977].

During 1952-1961, three companies operated commercial plants for the recovery of uranium from phosphoric acid. Solvent extraction was the main method of recovery but many operating difficulties were experienced. Recently an improved solvent extraction process was developed at the Oak Ridge National Laboratory [Hurst *et al*. 1972; Hurst & Crouse 1974; Hurst *et al*. 1977] and subsequently a number of companies constructed pilot plants, using this and similar processes, at commercial phosphoric acid plants in the USA to demonstrate the feasibility of the process [Ring 1975, 1977].

In 1974, Gulf Research and Development Corp. successfully operated a demonstration unit mounted on two enclosed truck trailers [McGinley & Facer 1975]. In 1975, Uranium Recovery Corp. completed construction of a central processing refinery at Mulberry, Florida, to recover uranium from a strip solution which was transported from an extraction module at the W.R. Grace & Co. phosphoric acid plant [MJ 1977c; McGinley & Facer 1975]. At full capacity this unit should yield 120 t of uranium concentrates per year but problems were encountered during start-up of the Davy Powergas extraction module, and extensive modifications were required [NW 1978]. Although economics are still uncertain, it seems unprofitable to construct modules at plants with a uranium production of less than 50 t/year or located more than about 80 km from the central refinery [McGinley & Facer 1975].

Freeport Chemical Co. have recently announced a $30 million project to recover uranium from a 750 000 t/year P_2O_5 phosphoric acid production plant in Louisiana [MJ 1977c]. Construction is scheduled for completion in 1978 with the plant reaching 85 per cent of rated capacity (260 t U/year) by mid-1979. Gardiner Inc. were also reported to have commenced construction of a $15 - 20 million recovery plant at East Tampa, Florida, with a capacity of 170 t U/year [CE 1978]. Union Carbide and Westinghouse Corp. have developed and tested solvent extraction processes, but have yet to reach decisions on commercial operations [MJ 1977c].

In other countries, PNC Corporation in Japan have announced plans to construct a pilot plant at Ningyo-Toge to test a solvent extraction process. As all phosphate rock processing in Japan is carried out in three or four plants which produce wet-process phosphoric acid, about 300 t U/year can probably be recovered economically [Wilson 1976]. A solvent extraction process has also been successfully tested in Spain, but the maximum production of uranium from this source is limited to 150 t U/year [Gomez *et al*. 1977]. In addition, the USSR is reported to have offered aid to Morocco for a feasibility study on extracting uranium from newly discovered Moroccan phosphate deposits [NW 1974].

10.2 Uranium from Copper Waste Dump Leaching Solutions

The leach solutions from many copper waste dumps in the western United States contain 1 - 10 mg U ℓ^{-1}. Although this is a very low concentration, solution flows are large enough (up to 2 $m^3 s^{-1}$) to yield a potential source of uranium of 800 - 900 t/year. Pilot plant tests by the US Bureau of Mines confirmed that the uranium was easily recovered by an ion exchange system [Facer 1973, 1977; McGinley & Facer 1975].

Recently, plans for the first commercial plant for the recovery of uranium from copper waste dump solutions were announced. A continuous countercurrent ion exchange system will be used to treat 0.5 $m^3 s^{-1}$ of copper leach solution containing 5 mg U ℓ^{-1} [EMJ 1977c].

11. PRODUCTION OF NUCLEAR PURITY PRODUCTS AT THE MINE SITE

In principle, the integration of milling and refining offers the potential of simplifying the overall process required for producing nuclear purity products from uranium ore. Hardy [1972] and Merritt [1971] reviewed the many approaches which have been studied for producing high purity products at the mine site. The only processes which continue to have been studied since then are those directed towards the production of nuclear purity uranium tetrafluoride (UF_4), namely the PNC process [Takenaka & Kawate 1976] being developed by the Power Reactor and Nuclear Fuel Corporation in Japan and two similar processes being developed by Pechiney Ugine Kuhlman in France [Boutonnet 1975].

The PNC process involves chloride conversion and stripping in an amine solvent extraction process to prepare a uranyl chloride solution which is then electrolytically reduced to uranous chloride. The electrodes of the cell are separated by a cation exchange membrane which enables hydrogen ions (produced by dissociation of water at the anode) to pass through and take part in the reduction reaction [Iammartino 1975]. Above 90°C, $UF_4.3/4\ H_2O$ is precipitated by adding 50 per cent hydrofluoric acid and then separated by filtration, washed, dried and dehydrated at 350°C to UF_4 [Takenaka & Kawate 1976].

The PNC flowsheet is similar to the French SIMO process for the conversion of uranyl nitrate solution to UF_4 which involves conversion of the nitrate solution to uranyl sulphate, electrolytic reduction and precipitation of $UF_4.3/4\ H_2O$ as outlined for the PNC process. The French cell uses mercury cathodes and platinum anodes. Boutonnet [1975] described the possibility of extending this process to the production of

UF_4 at the mine site. In addition, he outlined a process using sulphur dioxide in the presence of a copper salt to chemically reduce a uranyl chloride solution for the precipitation of $UF_4.3/4\ H_2O$. The chemical reduction step is similar to the Winlo process [Merritt 1971] developed in the USA, and the uranyl chloride solution is produced by chloride conversion of uranyl sulphate as in the PNC process.

No plant scale application of these processes has been reported and, despite their technical feasibility, they still appear to suffer from the disadvantage described by Hardy [1972]. These include higher costs at remote mine sites for chemicals, power, labour (and infrastructure), a smaller scale of operation than large refineries, and the need to obtain a satisfactory additional margin for selling UF_4 rather than yellowcake to existing refineries which are designed for conversion of yellowcake to UF_4.

12. WASTE TREATMENT AND POLLUTION CONTROL

The environmental impact of the uranium fuel cycle including uranium mining and milling and the requirements for waste treatment and pollution control have attracted increased interest over the last five years [USAEC 1974c; USEPA 1973]. Table 4 summarises the various potential pollutants and measures for their control. Major emphasis is placed on treatment of liquid wastes, mine water and run-off, and retention of tailings to minimise possible leaching and emanation of radon [Beverly 1976; Bonhote 1976; Costello 1977; Davy 1976; Kennedy *et al*. 1977; Levins *et al*. 1977]. Holloway *et al*. [1975] produced an extensive bibliography on uranium tailings containing more than 1200 references. Sears *et al*. [1975] reviewed waste treatment alternatives in the milling of uranium ores and correlated costs of treatment with the corresponding reduction in environmental impact for use in establishing 'as low as practicable' guides. The US Nuclear Regulatory Commission is preparing a draft Generic Environmental Impact Statement on uranium milling in the US up to the year 2000; this is scheduled for publication late in 1978 [Martin & Miller 1978; NF 1977c].

In 1976, the International Atomic Energy Agency published a Code of Practice for the Management of Wastes from the Mining and Milling of Uranium and Thorium Ores [IAEA 1976b]. International conferences on radon [IAEA 1975] and radiation protection [ILO 1976] in uranium mining and milling were held in 1973 and 1974, respectively. A Code of Practice on Radiation Protection in the Mining and Milling of Radioactive Ores was issued in Australia [Commonwealth Department of Health 1975].

12.1 Liquid Wastes

IAEA [1976b] recommended minimising discharges of effluents to surface or underground waters by restricting the use of fresh water, maximising recycling of mine water and liquid wastes, and using these liquids to slurry solid tailings being discharged to the tailings retention system. In areas where the natural rate of evaporation exceeds the rainfall, the waste retention system may be designed so that no discharge of liquid wastes is required. Otherwise, the tailings should be settled before liquid is decanted for discharge. Suitable treatment procedures should be employed to reduce contaminants.

The main process for the treatment of raffinate-tailings slurry from the acid leaching process is neutralisation with lime or a combination of limestone and lime [IAEA 1976b; Levins *et al.* 1977; Ryan & Alfredson 1975, 1976]. Neutralisation raises the pH, reduces the concentrations of dissolved salts including heavy metal impurities, and removes amine by adsorption on the precipitated solids. It has been adopted for proposed mills in Australia [Pancontinental 1977; Queensland Mines 1977; Ranger 1974], but has not been widely practised in the USA. Ryan & Alfredson [1976] showed that limestone was more efficient than lime for neutralisation at < pH 4.5 and that a two-stage precipitation, using limestone in the first stage at approximately pH 4 and lime in the second, without interstage separation of solids, may be advantageous.

After neutralisation and clarification, further treatment to remove radium may be necessary by adding $BaCl_2$ in the presence of excess sulphates to coprecipitate barium-radium sulphate [Beverly 1976; Ryan & Alfredson 1975, 1976]. Clarification of the effluent before release is very important. Radium removal has been practised in Canada since 1965 [Yourt 1975]; however, certain difficulties have been encountered, apparently because the radium precipitate redissolves in the ponds [IAEA 1976b].

12.2 Tailings

Tailings are normally retained in an embankment retention system, the design of which depends on such factors as the throughput and characteristics of the ore, topography and availability of land, net evaporation rate, soil permeability and embankment construction materials [IAEA 1976b]. In the US, in particular, standards and design guides for construction of these dams have been developed [ANSI 1974; USAEC 1974d]. Stabilisation and rehabilitation of tailings are being increasingly investigated with the aims of eliminating dispersion, reducing radon emission and background radiation levels and minimising leaching which may lead to pollution of groundwaters [Beverly 1976; Kennedy *et al.* 1977]. Alternatively return of tailings to the open-cut is being seriously considered since it minimises the need for further surveillance [Scarano & Lineham 1978].

Several alternatives to conventional tailings dam construction have been proposed. Robinsky [1975] developed a method of discharging thickened tailings as a conical or hill-like deposit which eliminates the steep slopes of conventional dams. The resulting hill was claimed to be more stable than conventional tailings dams. The use of horizontal belt vacuum filters in French-designed uranium mills (see Section 3.5) has been claimed to eliminate the need for tailings dams as the tailings are in a compact form and can be back-filled into trenches or heaped and rehabilitated on a continuous basis.

Where filters are used rather than CCD systems, the cake is usually repulped with a minimum amount of water and pumped to the tailings dam as a slurry. Tielens [1977] suggested that some tailings have thixotropic characteristics and might be pumped as a sludge with economic benefits. This would then allow disposal of an essentially solid waste.

12.3 Extraction of Radium from Tailings

If radium-226 and its parent thorium-230 could be removed from the tailings and concentrated for special disposal, the tailings could be disposed of as normal mine tailings [IAEA 1976b]. Tests have shown that most of the radium-226 in the tailings can be removed by leaching with strong solutions of HNO_3 or HCl at elevated temperatures [Seeley 1977; Ryon *et al.* 1977; Borrowman & Brooks 1975; Ring & Rapisarda 1977]. Recently, Levins *et al.* [1978] reported that it is technically feasible to remove more than 90 per cent of the radium by washing the tailings in a strong solution of sodium chloride or nitrate at room temperature. Release of radium is rapid, occurring within minutes. Low pulp densities are required in all processes for maximum dissolution of radium which could then be precipitated as a mixed barium-radium sulphate. To be economically acceptable reagent costs require the leach solution to be recycled after removal of radium.

13. CONCLUSIONS

Although the basic acid and alkaline leaching flowsheets which were developed in the 1950s and 1960s are still widely used, there have been significant developments in the individual unit operations during the 1970s as described above. These developments will enable economic treatment of lower grade ores despite increased capital and operating costs and more stringent requirements for environmental protection. In Australia, at least initially, the proposed new large mills will use proven conventional technology to guarantee performance, rather than new developments not yet demonstrated on a large scale.

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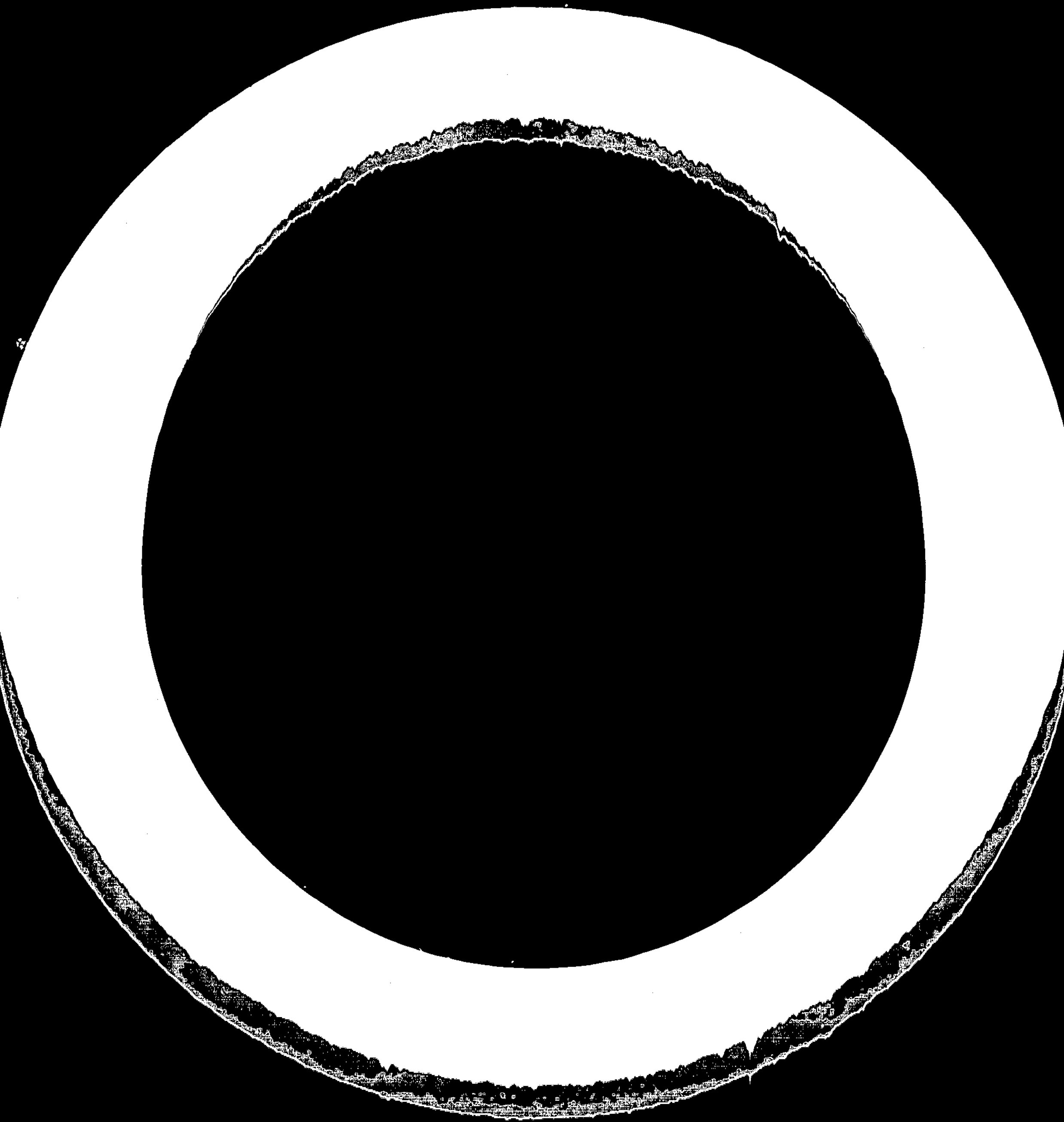

TABLE 1

AUTOGENOUS AND SEMI-AUTOGENOUS MILLING APPLICATIONS IN THE URANIUM INDUSTRY

Company, Plant	Description	Date of Operation	Reference
Canada			
Eldorado Nuclear Ltd. Beaverlodge	Wet, semi-autogenous, run-of-mine ore feed	1960s	Fish 1975; Lendrum & McCreedy 1976
Gulf Minerals Canada Ltd. Rabbit Lake	Wet, semi-autogenous, run-of-mine ore feed	1975	IAEA 1976a; Lendrum & McCreedy 1976
Niger			
Somair, Arlit	Dry autogenous cascade, feed primary crushed	1971	IAEA 1976a
Consortium, Akouta Niger Govt., etc.,	Dry autogenous milling	1978	IAEA 1976a
United States			
Union Carbide Corp. Uravan	Dry autogenous Aerofall mill, feed primary crushed	1960s	Jones 1977; Merritt 1971
Lucky Mac Uranium Corp. Shirley Basin	Wet, semi-autogenous cascade, run-of-mine ore feed	1971	Jones 1977; USAEC 1974a
Western Nuclear Inc. Jeffrey City	Wet, semi-autogenous cascade, run-of-mine ore feed	1976 (Plant expansion)	Jones 1977; McManus 1977; Seidel 1976
Sohio Oil, Reserve Oil and Minerals, Laguna	Semi-autogenous cascade	1976	Jones 1977; McManus 1977
United Nuclear, Church Rock	Semi-autogenous mill	1977	Jones 1977; McManus 1977
The Anaconda Co., Bluewater	Autogenous milling	1978 (Rebuilding plant)	Jones 1977; WM 1977a
Rocky Mountain Energy, Shirley Basin	Semi-autogenous mill	1978	Jones 1977

TABLE 2

SOLUTION MINING OPERATIONS

Company	Location	Design Capacity t U/year	Operation Commenced	References
Utah Construction & Mining Co.	Shirley Basin		1968	Anderson & Ritchie 1968
US Steel, Niagara Mohawk	Clay West in Live Oak County near George West, Texas	100 +	1975	Davis *et al*. 1976 Narayan & Rand 1976, EMJ 1977a Facer 1977, White 1975
US Steel	Burns Ranch, Texas	60	1976	EMJ 1977a, Facer 1977
Wyoming Mineral Corp.	Bruni, Texas	40 *	1977	EMJ 1977a, Facer 1977
Wyoming Mineral Corp.	Ray Point, Texas	200	1977	EMJ 1977a, Facer 1977
Wyoming Mineral Corp.	Irigaray Ranch Power River Basin, Wyoming	200	1977	OECD/IAEA 1977, NW 1977
Union Carbide	Duval County, Texas		1976	EMJ 1977a, OECD/IAEA 1977
Intercontinental Energy	Pawnee, Texas	70	1977	EMJ 1977a, Facer 1977
Intercontinental Energy	Ray Point, Texas	120	1978	EMJ 1977a

+ Expanding to 400 t U/year [EMJ 1977a; Davis *et al*. 1976]

* Expanding to 100 t U/year

TABLE 3

COMPARISON OF SPECIFICATIONS FOR YELLOWCAKE

Constituent (wt %)	Allied Chemical Corp.		Eldorado	Kerr McGee Corp.	Comhurex	British Nuclear Fuels Ltd
	Standard Concentrates (a)	Maximum Limit Concentrates (a)	Limit Without Surcharge (b)	Limit Without Surcharge (c)	Maximum Limit (d)	Limit Without Surcharge (e)
Uranium, U	75 min	65 min	60 min	65 min	60.0 min	50.9 min
Moisture	2.00	5.00	5.00	7.50	5	10
Carbonate, CO_3	0.20	0.50#	2.00	2.00	2.00	2.00
Sulphate, SO_4	3.00	12.00	10.50	10.50	10.50	10
Vanadium, V_2O_5	0.10	0.75	0.18	0.18	1.80	1.00
Phosphorus, PO_4	0.10	1.00	0.52	0.52	1.10	2.00 + 1.31 Fe+
Fluorine, F	0.01	0.10	0.15	0.15	0.15	x
Halogens, Cl+I+Br	0.05	0.10	0.25*	0.25	0.25	0.50x
Arsenic, As	0.05	0.10	1.00	1.00	1.00	2.00
Boron, B	0.005	0.10	0.15	0.15	0.15	0.20
Calcium, Ca	0.05	1.00	1.00	1.00	1.15	-
Iron, Fe	0.15	1.00	-	1.50	-	-
Magnesium, Mg	0.02	0.50	-	1.0	-	-
Molybdenum, Mo	0.10	0.30	0.15	0.15	0.45	0.60
Potassium, K	0.20	3.00	-	-	-	-
Silicon, SiO_2	0.50	2.50	-	2.1	-	4.00
Sodium, Na	0.50	7.50	-	-	-	-
Thorium, Th	-	-	2.00	2.00	-	-
Titanium, Ti	0.01	0.05	-	-	-	-
Zirconium, Zr	-	-	-	0.5	-	-
HNO_3 Insoluble	-	-	0.1	0.1	0.1	-
Extractable Organic Material	-	-	0.1	0.1	-	-

\# Applies only for >0.5% Na
* As chloride
\+ Lesser of 6.00 and 2.00 + 1.31 Fe
x Includes F

Basis of specification
(a) Dry weight (Allied Chemical undated)
(b) Contained U except for U itself on as sampled basis (Eldorado undated)
(c) Concentrates as received (Kerr McGee undated)
(d) Contained U except for U itself on a dry basis (Comhurex undated)
(e) Contained U_3O_8 except for U itself on a dry basis (BNFL undated)

TABLE 4

POTENTIAL POLLUTANTS AND THEIR CONTROL

SOURCES	POLLUTANTS	TREATMENT AND POLLUTION CONTROL MEASURES
MINING		
a Top soil	Suspended solids in run-off, dust	Control drainage, stabilise by landscaping and revegetation.
b Overburden Below-ore grade material (bogum)	Suspended solids, dissolved heavy metals (e.g. Cu, Zn, Pb, Cd), uranium and radium arising from bacteriological and chemical leaching by rainwater, dust	Construct bunds around heaps to collect seepage and run-off. Depending on concentration of pollutants, treat to remove radium and heavy metals or release to surface waters at times when adequate dilution is ensured. Stabilise by landscaping and revegetation or return chemically active bogum to mine.
c Mine water Ore stockpile run-off	Generally higher concentrations of uranium, radium, radon and heavy metals than (b)	Collect and use as feed water to mill. Excess impounded, evaporated or treated before discharge. Disposal of pollutants in tailings dam.
d Blasting, drilling, mining, wind erosion	Radon, silica and general dusts	Forced ventilation in underground mines and possibly small, high grade opencuts. Spray dusty areas with water.
MILLING		
e Tailings	Heavy metals, uranium, radium in seepage	High standard of design and containment. Seal embankments with impervious clay to limit seepage, collect and pump back seepage.
	Radon release, airborne tailings and dust	Maintain tailings under water during mill operation. Return to mine or cover with soil and revegetate after mill closure.
f Raffinate, other process liquid wastes	High concentrations of dissolved salts, heavy metals and acidity.	Neutralise with lime and/or limestone. Impound in tailings dam. Recycle from tailings dam to minimise liquid waste volume.
g Crushing, screening	Radioactive dusts, silica, radon	Collect dusts on filters or in wet scrubbers, disperse radon.
h Acid production	Oxides or sulphur, sulphuric acid mist	Demisters, scrubbers. Tall stack for dispersion.
i Calcining and Packaging	Uranium oxide (U_3O_8)	Airhoods, ventilation, filters, wet scrubbers, face masks.

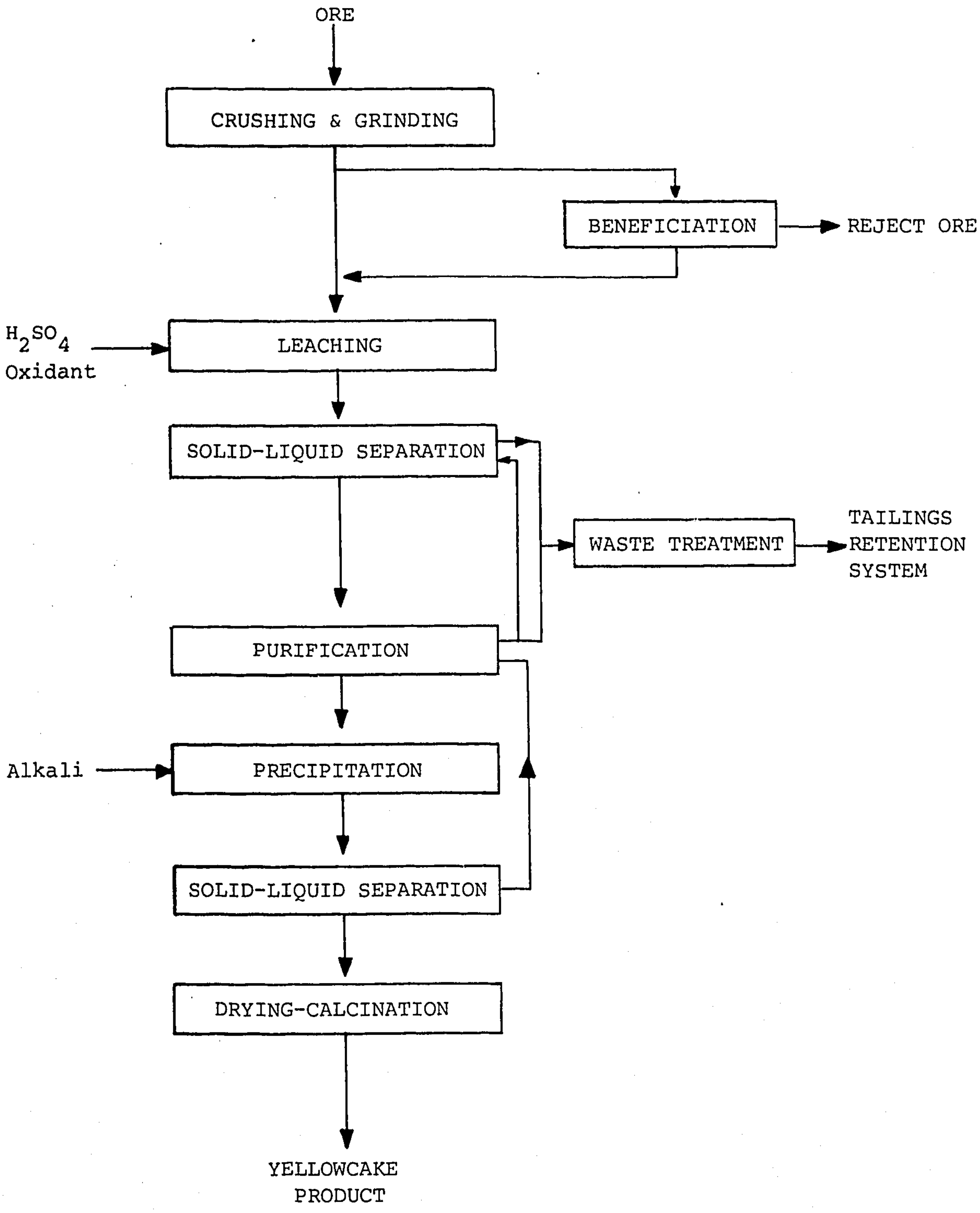

FIGURE 1. ACID LEACHING FLOWSHEET

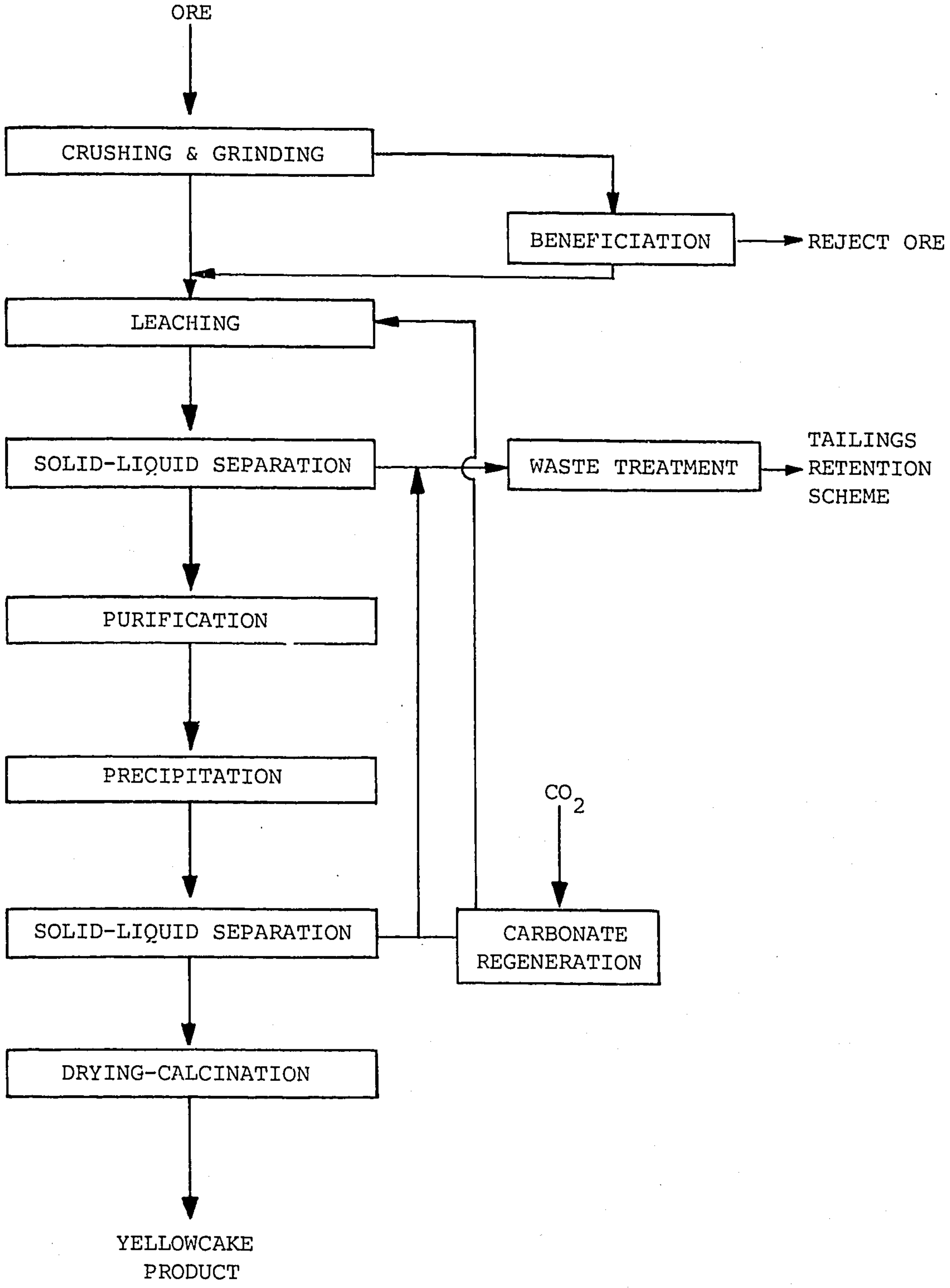

FIGURE 2. ALKALINE LEACHING FLOWSHEET

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