

# Utilization of Seawater Brines for the Production of High Purity Magnesium Oxide and Magnesium Hydroxide

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## ABSTRACT

Residual seawater brines from sodium chloride plants and potash waste brines are rich sources of magnesium and other elements, but the impurities present in them make their exploitation difficult. It is demonstrated here that magnesium oxide and magnesium hydroxide of high purity can be produced from these sources.

In the developed method for the production of magnesium oxide, the desulfated brine from a potassium-magnesium-sulfate, double salt, pilot plant is solar-concentrated to 1.31 specific gravity to reduce its sodium chloride concentration. This brine is evaporated to dryness. The dry salt is decomposed and converted to magnesium oxide by calcination at about 1500°C.

For the production of magnesium hydroxide, the brine from the sodium chloride plants is concentrated to 1.31 specific gravity and then mixed with ammonia. The magnesium hydroxide thus formed is separated from the liquor by filtration, washed with water and dried. The ammonia is recovered by distillation and recycled to the process.

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## INTRODUCTION

Magnesium oxide is commonly obtained by burning magnesium hydroxide. Magnesium hydroxide in turn is produced from magnesium soluble salts by reaction with calcined dolomite or calcined limestone (Chesny, 1936; Manning, 1938, 1947; Shreve, 1956; Vohra, et al. 1968; Gormly, 1969; Hall and Spencer, 1969; Hall and Spencer, 1973; and Fernandez-Lozano and Richardson, 1975). Magnesium oxide may also be obtained from dead burning naturally occurring magnesites or from magnesium chloride-rich brines (Norden, 1956; Hall and Spencer, 1973; and Norden, 1974). At the present time the magnesium hydroxide is produced from seawater using calcined dolomite or limestone as a magnesium precipitant agent. The seawater is the universally used raw material, but due to its low content of magnesium, it is necessary to process large quantities of it, which causes the fixed costs and operation costs to be high. In addition, it is necessary to have available high purity dolomite or limestone.

Sources richer in magnesium than seawater are the residual brines from seawater desalination and sodium chloride plants. However because of the high sulfate ion concentration, which in the presence of  $\text{Ca}(\text{OH})_2$  precipitates as  $\text{CaSO}_4 \cdot 2\text{H}_2\text{O}$  together with  $\text{Mg}(\text{OH})_2$ , the separation of both products being very difficult and expensive, these sources have been seldom used for the production of  $\text{Mg}(\text{OH})_2$  or  $\text{MgO}$ . It is demonstrated here that these brines may be used for the production of high purity  $\text{Mg}(\text{OH})_2$  by the ammoniation process proposed in this work.

Other rich and purer sources of magnesium are the residual brines from the plants that produce potassium-magnesium-sulfate double salts (SPM salts) from seawater brine and the brines from Epsomite-Carnallite-Sylvite plants (Fernandez-Lozano, 1973, 1974, 1975 and 1976). These brines may be used for the production of high purity  $\text{MgO}$ .

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by the direct calcination process proposed in this work or for the production of  $\text{Mg}(\text{OH})_2$  by the ammoniation process.

In Israel, high purity  $\text{MgO}$  is being produced from purified Dead Sea brines by decomposition of the chlorides at high temperature (Norden, 1974). A similar process was used in a plant constructed in the U.S.A. and placed in operation in the summer of 1953, but in this case the raw material was rich purified  $\text{MgCl}_2$  brine from a Langbeinite plant (Norden, 1956; Gloss and Dancy, 1958). The process was dismantled because of the lack of market for the  $\text{HCl}$  byproduct. In 1957, a U.S. Government patent was issued which outlined a process for purifying magnesium chloride rich brines (Dancy, 1957). This patent, even though reasonably informative, does not allow fair conclusions to be reached on the possibilities of using seawater brines as raw materials in the production of high purity  $\text{MgO}$ , neither does it provide any information on rates.

No prior studies have been published on the production of magnesium oxide and magnesium hydroxide by the proposed methods from concentrated seawater brines. It was this lack of information and the importance of the subject which prompted this investigation.

Figure 1 shows the quantity of different chemicals that could be recovered from the brine of a one million ton  $\text{NaCl}$  plant from seawater. If all these products were sold at the present world price, the value would be quite large.

### SCOPE OF THE INVESTIGATION

The main objectives of the present investigation were:

1. To develop and demonstrate in the laboratory two simple processes for the production of high purity magnesium oxide and magnesium hydroxide from concentrated seawater brines.
2. To prepare suitable seawater brines for the processes.
3. To prepare a crude magnesium oxide suitable for the production of high purity magnesium oxide.
4. To evaluate some of the most important variables that control the impurities and density of the  $\text{MgO}$  product. These variables are a) Type of reducing agents used, b) the concentration of reducing agent, c) calcination temperature, and d) calcination time.
5. To evaluate some of the most important variables that control the recovery efficiency and purity of the  $\text{Mg}(\text{OH})_2$  product. These variables are a) Brine concentration, b) ammonia concentration, c) reaction temperature, and d) reaction time.
6. To present an outline of the proposed processes.

These six objectives have been met by the investigation summarized in this paper.

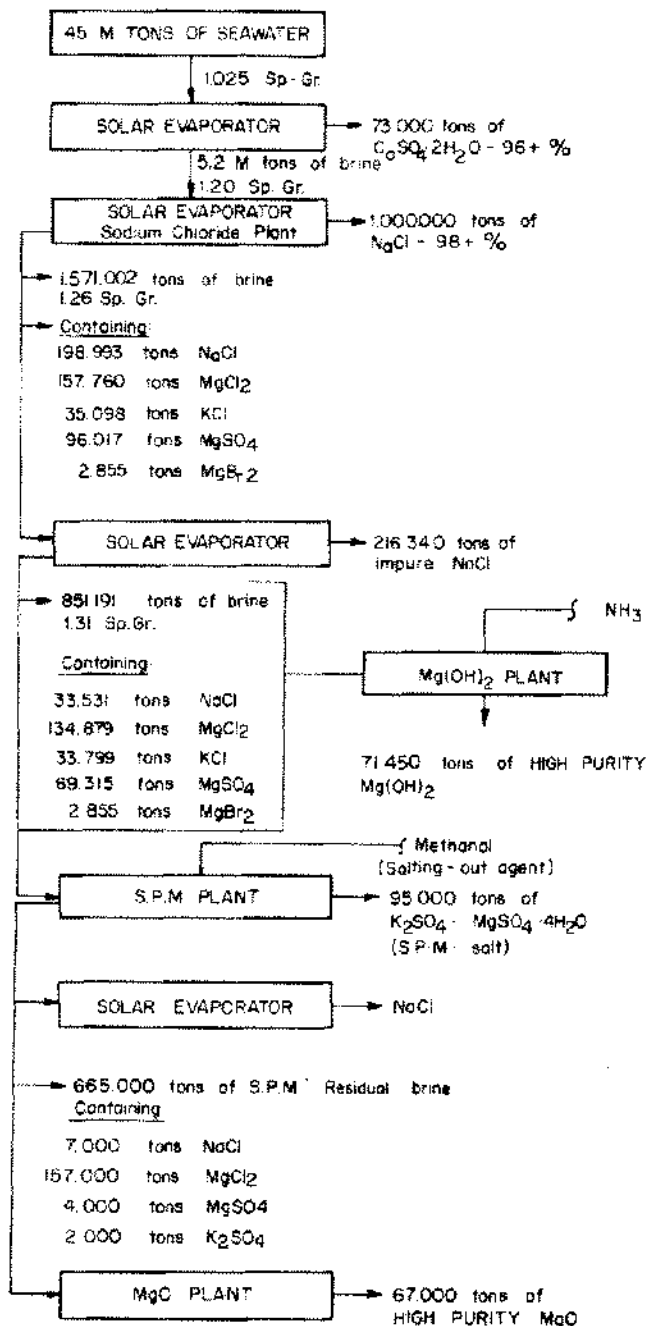


Figure 1. The seawater brine from Halite Plants is an important source of many chemicals.

### DESCRIPTION OF THE MAIN SETS OF APPARATUS

The apparatus employed in the preparation of  $\text{MgO}$  was designed to accomplish high temperature calcination of small samples. Provision was made for setting and maintaining a constant temperature, for simulating somewhat the

atmosphere of a direct fire dryer and for fairly rapid movement of combustion gases and volatilized impurities. This apparatus consisted of the following essential parts: 1) a cylindrical resistance furnace with an adapted gas burner. The 30 cm  $\times$  10 cm cylindrical heating chamber is an enclosure with a refractory lining, a surrounding layer of heat insulation and an outer steel jacket; 2) three thermocouples; 3) a gas burner; 4) a potentiometer, and 5) three high temperature calcination cups. The chamber was connected through a silicon carbide tube to a falling water line; this procedure facilitates rapid evacuation of the combustion and volatilized gases from the heating chamber. A schematic diagram of the assembled apparatus is shown in Figure 2.

The apparatus employed in the preparation of  $Mg(OH)_2$  was designed to accomplish the rapid mixing of measured amounts of brine and ammonia and the rapid separation of the resulting crystalline precipitate of  $Mg(OH)_2$  from the reaction solution. Provision was made for setting and maintaining a constant temperature, for agitation and for measuring the time of the reaction. The reactor proper was a Q.V.F. internally cooled pressure vessel with internal baffles, a sloping bottom, and a center outlet tube in the bottom. The reactor could be pressurized to 50 psig for rapid filtration of its contents at the proper moment. The filter media was a disk of medium porosity canvas cloth placed on a perforated stainless steel plate as shown in Figure 3. The injector was a jacketed pressure vessel with a solenoid valve in the discharging port for rapidly charging the solution into the reactor. The temperature in the reactor was measured by a chromel-alumel thermocouple, the reaction times were measured with an electronic timer and the ammonia concentration in solution was determined by analysis. A schematic diagram of the assembled apparatus is given in Figure 3.

## LABORATORY WORK

**Preparation of residual SPM seawater brine.** The residual brine from a salt field with 1.26 specific gravity was solar concentrated to 1.31 specific gravity in plastic rectangular tanks. The evaporation proceeded under natural conditions, with the temperature of the brine varying between 26° and 45°C. The brine was then separated from the salts by vacuum filtration. The concentrated brine, free of solids, was discharged into an agitated vessel and methanol added while stirring. The quantity of methanol added was equal to 70% by weight of the brine-alcohol mixture. The residence time in the methanulator was adjusted to about 35 minutes and then the slurry was discharged into a vacuum filter, where the solids ( $K_2SO_4 \cdot MgSO_4 \cdot 4H_2O$ ) were removed and discarded and the filtrates were distilled for recovering the methanol. The brine, free of methanol, was solar concentrated again to 1.31 specific gravity to remove some more NaCl from solution. The results shown in Table 1,

TABLE I  
Composition of the Residual Brine  
from an SPM Plant

Components	Composition, % Weight
$MgCl_2$	23.625
NaCl	1.051
$MgSO_4$	0.915
KCl	0.466
$MgBr_2$	0.181
$B_2O_3$	0.026
$CaSO_4$	0.062
$SiO_2$	0.023
$Fe_2O_3$	0.003
$Al_2O_3$	0.002
MnO	0.002
$H_2O$	73.644

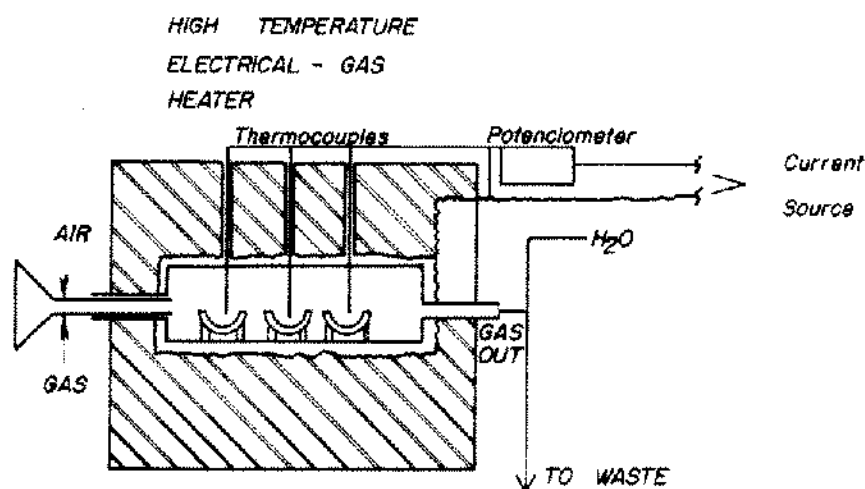
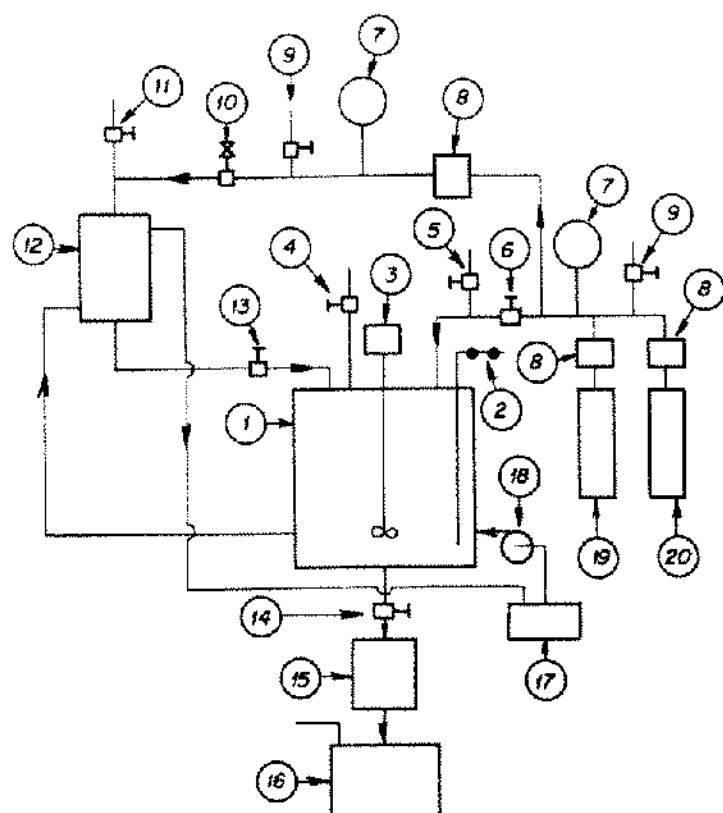


Figure 2. Schematic diagram of the apparatus used in the calcination work.

## LEGEND



1. REACTOR
2. THERMOCUPLE
3. STIRRER AND STIRRER MOTOR
4. REACTOR CHARGE PORT
5. REACTOR VENT PORT
6. REACTOR PRESSURIZATION AND AMMONIA INJECTION VALVE
7. PRESSURE GAGES
8. PRESSURE REGULATORS
9. VENT PORTS
10. INJECTOR PRESSURIZATION VALVE
11. INJECTOR CHARGE PORT
12. INJECTOR
13. INJECTOR DISCHARGE PORT
14. REACTOR DISCHARGE PORT
15. FILTER
16. FILTRATE RECEIVER
17. CONSTANT TEMP-BATH
18. WATER CIRCULATING PUMP
19. GAS BOTTLE
20. AMMONIA GAS BOTTLE

Figure 3. Schematic flow diagram of the ammoniator reactor crystallizer.

reveal that the brine so prepared is very pure and rich in magnesium chloride. The ionic composition of the liquid and solid phases was chemically evaluated by the procedure proposed by Rafols (1969).

**Preparation of crude magnesium oxide.** The purified brine, free of most of the potassium, sulfates, calcium, carbonates and sodium was evaporated to dryness at 90°C for three hours in a thermo-circulator low temperature oven and analyzed; the results are summarized in Table 2. Weighed

samples of 5 grams of the dry salt were placed in the calcination cups and calcined in the apparatus shown in Figure 2.

The test was begun by heating the samples to the test temperature, keeping the test temperature for the desired calcination time. Finally, the fire was stopped, and the sample cooled, weighed and analyzed. The results are summarized in Table 3 and reveal that the conversion of magnesium chloride to magnesium oxide is almost complete in 40 minutes at 500°C. The results reveal also that very

TABLE 2

Composition of the Salts Obtained by  
Evaporation to Dryness of the Residual Brine  
from an SPM Plant

Components	Composition, % Weight
MgCl <sub>2</sub>	43.995
NaCl	1.955
MgSO <sub>4</sub>	1.701
KCl	0.868
MgBr <sub>2</sub>	0.377
B <sub>2</sub> O <sub>3</sub>	0.091
CaSO <sub>4</sub>	0.115
SiO <sub>2</sub>	0.040
Fe <sub>2</sub> O <sub>3</sub>	0.004
Al <sub>2</sub> O <sub>3</sub>	0.004
MnO	0.004
H <sub>2</sub> O	50.846

TABLE 3

Composition of the Crude Magnesium Oxide Obtained by  
Calcination at 500°C of the Dry Salt with a  
Composition as Shown in Table 2

Components	Composition, % Weight
MgO	78.980
NaCl	7.523
MgSO <sub>4</sub>	5.819
KCl	3.701
MgCl <sub>2</sub>	3.226
B <sub>2</sub> O <sub>3</sub>	0.102
CaSO <sub>4</sub>	0.427
SiO <sub>2</sub>	0.164
Fe <sub>2</sub> O <sub>3</sub>	0.024
Al <sub>2</sub> O <sub>3</sub>	0.016
MnO	0.016

little magnesium sulfate is decomposed under these conditions. It is important to note here that some magnesium chloride should remain in the crude magnesium oxide since it is necessary for the formation of magnesium oxychloride which acts as cementing bonding material, and improves pellets strength. Tests were conducted at other temperatures and heating times, but the results are not reproduced here for the sake of brevity.

**Binding promoters, reducing and sintering compounds.** Water is added to the mixture of the crude magnesium oxide and carbonaceous material to promote the formation of binding compounds. The water reacts with the magnesium oxide and undecomposed magnesium chloride to form magnesium oxychloride which is a cementing bonding material. This compound acts as binder when the crude magnesium oxide is briquetted. The quantity of water utilized in this study was equal to four percent by weight of the mixture of crude magnesium oxide and carbonaceous materials. The effects of this variable were not investigated because of the time limit of this study.

Carbonaceous materials act as reducing agents and were added to the crude magnesium oxide to assist in the volatilization of impurities during high temperature calcination. Several carbonaceous materials were tested and the results

are summarized in Table 5. These data reveal that any of the compounds tested were effective in the removal of impurities. Coal coke was used in this study since it removes impurities somewhat more efficiently than the other carbonaceous materials, and the quantities used were 1, 2, and 3 percent based on the weight of the crude magnesium oxide. The results are shown in Figures 4 and 5 and discussed later in this study.

Sintering compounds are added to produce magnesium oxide for basic magnesia bricks. These compounds belong to the silica family and consist of finely divided calcium oxide, talc, quartz, diatomaceous earth and silicon oxide. The amount added depends on the desired magnesium oxide content in the final produce and should be mixed with the crude magnesium oxide before the addition of water and after the addition of the carbonaceous material. These sintering compounds improve the hardness of the briquettes and have considerable effects on the ultimate chemical and physical properties of the final product.

**Pelletization.** To ensure intimate contact between the carbonaceous material and the impurities, the mixture of the crude magnesium oxide and carbonaceous material was briquetted before final calcination. It was found in the course of this study that calcination without briquetting results in inadequate removal of the impurities. The results are now shown here for the sake of brevity.

Pelletization of the crude magnesium oxide and carbonaceous material is particularly important with respect to the calcination operation. It facilitates the movement of the combustion gases and air in the calciner and minimizes the dust problems.

The pelletization can be made in any suitable apparatus. As stated before, the addition of water to the mixture of magnesium oxide and carbonaceous material promotes formation of cementing bonding materials, therefore improving the strength of the pellets. The pellets used in this study were small cylinders 2 mm  $\times$  5 mm formed at a pressure of 5,000 pounds per square inch.

**Preparation of high purity magnesium oxide.** The crude magnesium oxide was intimately mixed with the car-

TABLE 4

Composition of Magnesium Oxide Obtained by Calcination at 1500°C of the Crude Magnesium Oxide with a Composition as Shown in Table 3

Components	Composition, % Weight
MgO	99.174
SiO <sub>2</sub>	0.168
CaO	0.511
Fe <sub>2</sub> O <sub>3</sub>	0.025
Al <sub>2</sub> O <sub>3</sub>	0.018
MnO	0.018
SO <sub>4</sub>	0.044
Cl	0.040
Na	0.008
B <sub>2</sub> O <sub>3</sub>	0.008
K	0.004

TABLE 5

Composition of the Magnesium Oxide Produced by Calcination of Crude Magnesium Oxide with Different Carbonaceous Materials as Reducing Agents.

	Calcination time, 1 hour.	Calcination Temperature °C	Reducing Agent	% Wt.			
				Na	K	SO <sub>4</sub>	B <sub>2</sub> O <sub>3</sub>
Crude Magnesium Oxide				2.958	1.735	4.641	0.101
Reducing Agents							
Coal Coke	1500		3.0	0.004	0.004	0.009	0.007
Coal Tar Pitch	1500		3.0	0.005	0.004	0.012	0.008
Petroleum Coke	1500		3.0	0.006	0.005	0.009	0.009
Asphalt	1500		3.0	0.007	0.006	0.021	0.007

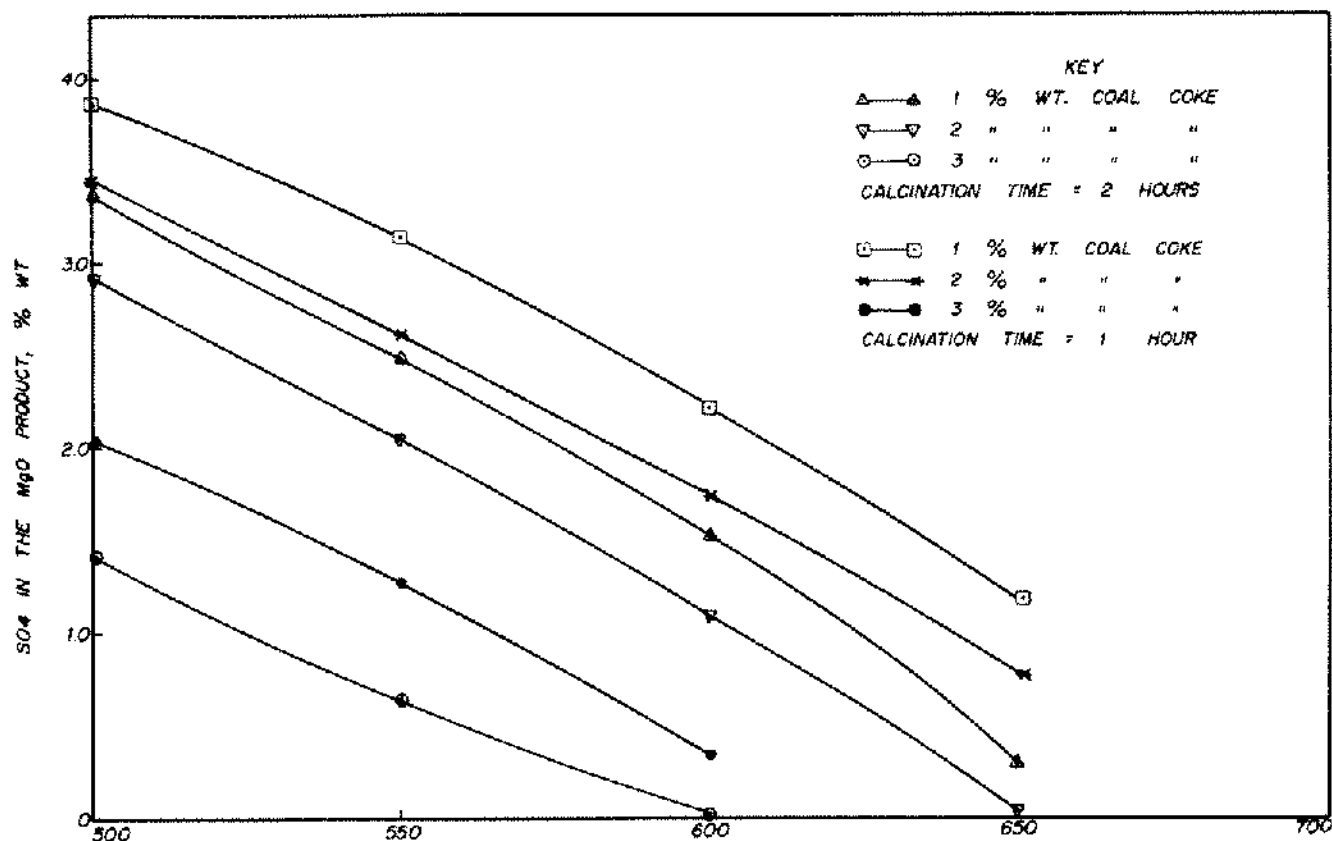


Figure 4. Effect of calcination temperature on the  $\text{SO}_4$  concentration in the  $\text{MgO}$  product.

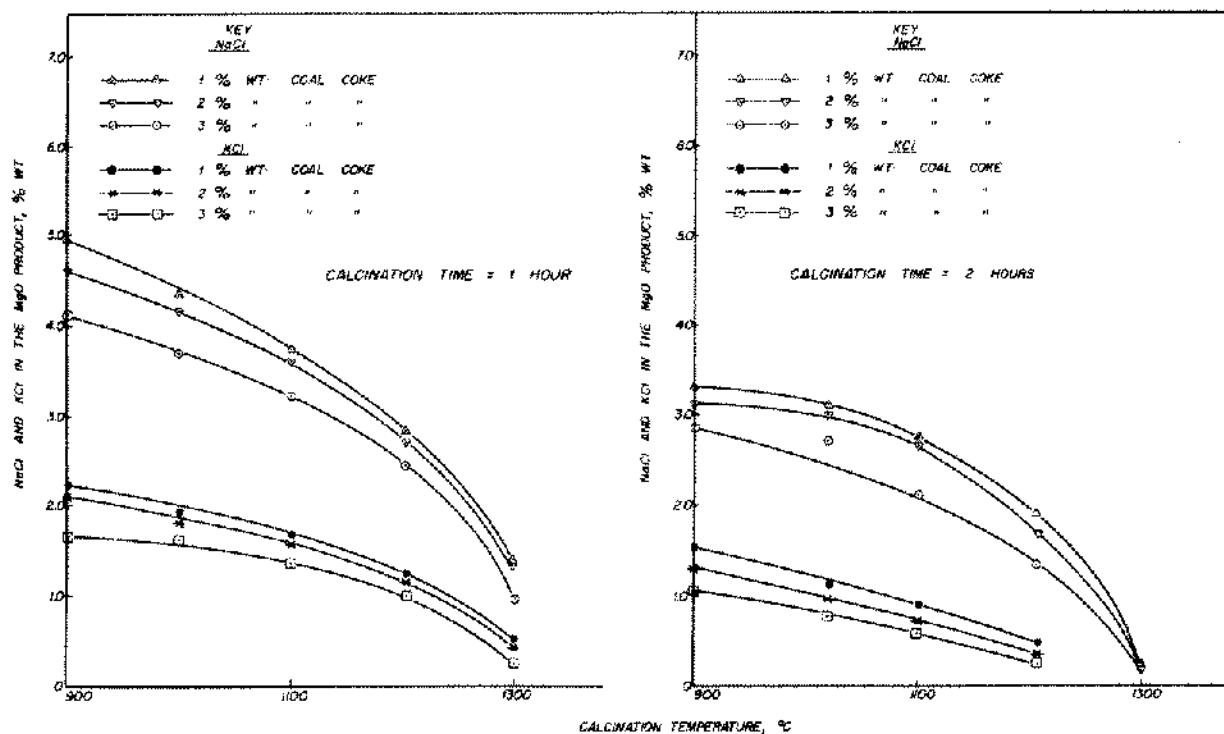


Figure 5. Effect of calcination temperature on the NaCl and KCl concentrations in the  $\text{MgO}$  product.

bonaceous material and water, then the mixture was pelletized as mentioned before in this study. Weighed samples of the crude magnesium oxide pellets were placed in the calcination cups and calcined in the apparatus shown in Figure 2. The calcination procedure followed in these tests was the same as that used in the preparation of the crude magnesium oxide. The results are summarized in Figures 4 and 5.

Figure 4 reveals that the concentration of sulfate ( $\text{SO}_4$ ) in the calcined product decreases almost linearly as temperature increases. The results also show that the ( $\text{SO}_4$ ) concentration decreases significantly as the percentage of coke in the crude magnesium oxide and calcination time increase. It is important to note here, that for a calcination time of two hours and 3 percent coke, the ( $\text{SO}_4$ ) concentration in the product is nearly zero at  $600^\circ\text{C}$ . The same results are observed at  $650^\circ\text{C}$  for a calcination time of 1 hour with 3 percent coke.

Figure 5 reveals that the  $\text{NaCl}$  and  $\text{KCl}$  concentrations in the calcined product decrease significantly as calcination temperature and calcination time increase. High coke concentration favors the removal of impurities, but its effectiveness is lower for these impurities than for  $\text{SO}_4$ . It is important to note here that for a calcination time of two hours and 3 percent coke, the removal of  $\text{NaCl}$  and  $\text{KCl}$  is almost complete at  $1300^\circ\text{C}$ .

The specific gravity of the calcine produced increases with calcination temperature from about 2.75 at  $500^\circ\text{C}$  to

3.25 at  $1500^\circ\text{C}$ . Higher coke concentrations in the crude magnesium oxide decrease the density of the product for the same calcination temperature, but this effect is lower at temperatures above  $1100^\circ\text{C}$ .

#### Preparation of high purity magnesium hydroxide.

The test was begun by charging 300 ml of seawater brine to the constant temperature ammoniator reactor crystallizer (Figure 3). A minimum of 20 minutes was allowed for thermal equilibrium to be established.

The crystallization was started by adding the ammonia to the brine in the crystallizer. The ammonia was added as ammonia gas from the ammonia bottle or as ammonium hydroxide from the injector. The stirrer control and timer were actuated automatically as the ammonia injection valve was opened. At any desired sampling time the reactor outlet valve was opened and the reactor contents discharged through the filter which retained the crystals. Air was blown through the crystals for a few seconds to remove some of the adhering solute and then the crystals were thoroughly washed on the filter with water. Finally the crystals were removed from the filter, dried, weighed and analyzed. X-ray diffraction analysis of samples washed with alcohol reveal the presence of  $\text{MgSO}_4 \cdot 6\text{H}_2\text{O}$ ,  $\text{MgCl}_2 \cdot \text{NH}_4 \cdot 6\text{H}_2\text{O}$  and  $(\text{K}, \text{NH}_4)_2 \text{Na}(\text{SO}_4)_2$  in addition to  $\text{Mg}(\text{OH})_2$ .

The effect of temperature on the kinetics of  $\text{Mg}(\text{OH})_2$  crystallization from 1.31 specific gravity seawater brine with ammonia was determined at  $30^\circ$  and  $20^\circ\text{C}$ . Typical results, as summarized in Figure 6, indicate a very rapid

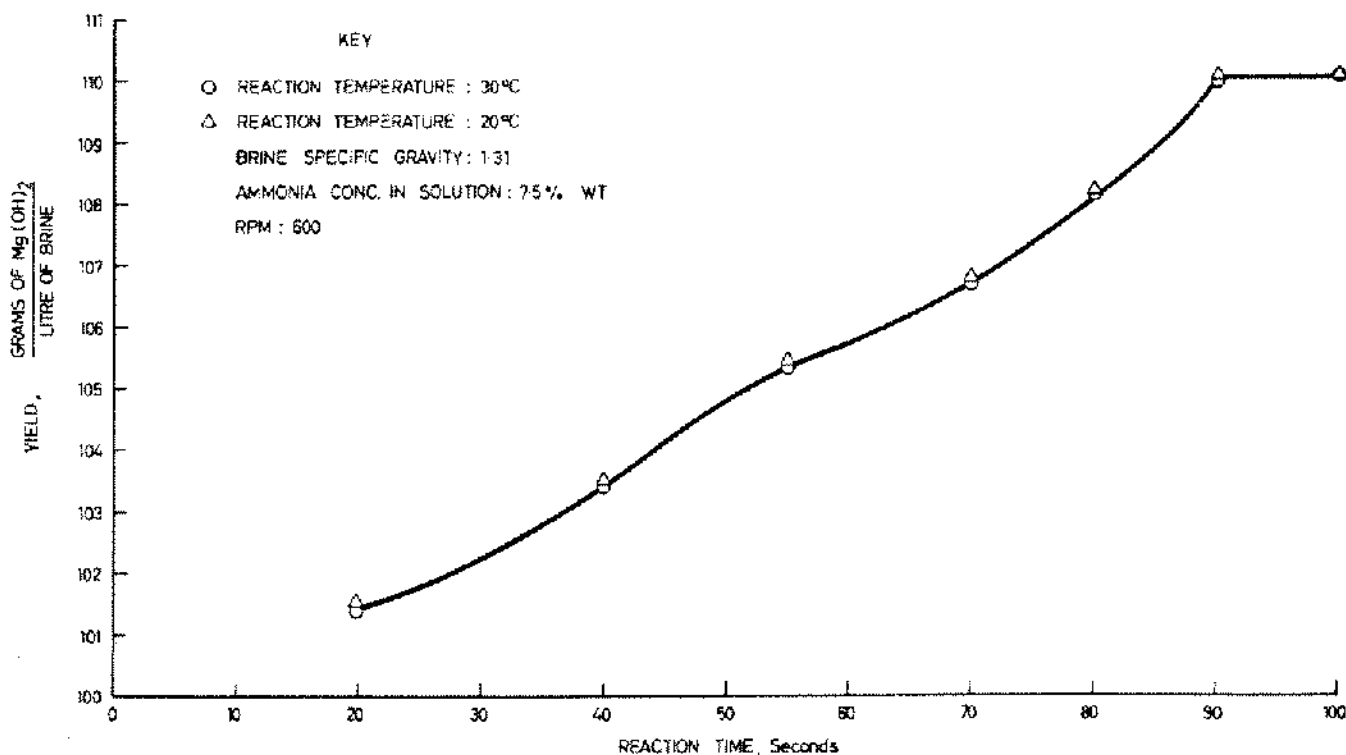


Figure 6. Effect of temperature on the kinetics of  $\text{Mg}(\text{OH})_2$  crystallization from seawater brine with ammonia.

TABLE 6

Amount of  $\text{Mg}(\text{OH})_2$  Recovered Per Litre of Brine and Purity as a Function of Brine Density.

Brine Density gm/cm <sup>3</sup>	$\text{Mg}(\text{OH})_2$ Recovered gm/lit	Purity of $\text{Mg}(\text{OH})_2$ % Wt.
1.261	41.60	99.88
1.283	80.45	99.96
1.310	110.10	99.96

Ammonia in solution = 7.6% Wt; reaction time = 60 secs.  
Reaction temperature = 30°C; RPM = 600

rate of crystallization, reaching maximum yield in about 90 seconds. The same figure shows that yield is almost independent of reaction temperature.

The tests were repeated with brine of different densities, all taking place at 30°C and with ammonia added in stoichiometric amounts to react with all the magnesium present. The results summarized in Table 6 reveal that the  $\text{Mg}(\text{OH})_2$  yield increases from 41.60 grams per liter for brine with a specific gravity of 1.26 to 110.10 for brine with a specific gravity of 1.31.

Tests were conducted using different amounts of ammonia with brine of 1.31 specific gravity, at 30°C. The results presented in Table 7 indicate that the maximum recovery efficiency corresponds to the ammonia concentration in solution being about 7.6% by weight.

Stirring rate data are not reproduced here for the sake of brevity, but a point is worth noting: the magnesium hydrox-

TABLE 7

Amount of  $\text{Mg}(\text{OH})_2$  Recovered Per Liter of Brine and Purity as a Function of Ammonia in Solution.

Ammonia in Solution % Wt.	$\text{Mg}(\text{OH})_2$ Recovered gm/lit	Purity of $\text{Mg}(\text{OH})_2$ % Wt.
2.54	9.45	99.5
4.95	31.10	99.6
7.25	104.64	99.8
7.62	110.10	99.6
7.99	109.05	99.6

Brine density = 1.31; reaction time = 90 seconds  
Reaction temperature = 30°C; RPM = 600

ide yield increased as the stirring speed increased up to about 600 rpm. Beyond this point, the recovery was independent of rpm.

### PROPOSED PROCESSES

#### Description of the process for the production of $\text{MgO}$ .

Examining the data presented in this study and cited literature, it is quite evident that residual brines from SPM and Epsomite-carnallite-sylvite plants are suitable raw materials for the production of high purity magnesium oxide. These results offer enough incentive to utilize and incorporate them in a magnesium oxide recovery process.

The schematic flow diagram of the proposed process is shown in Figure 7. The process is divided into three major operations, which in turn, are divided into various steps.

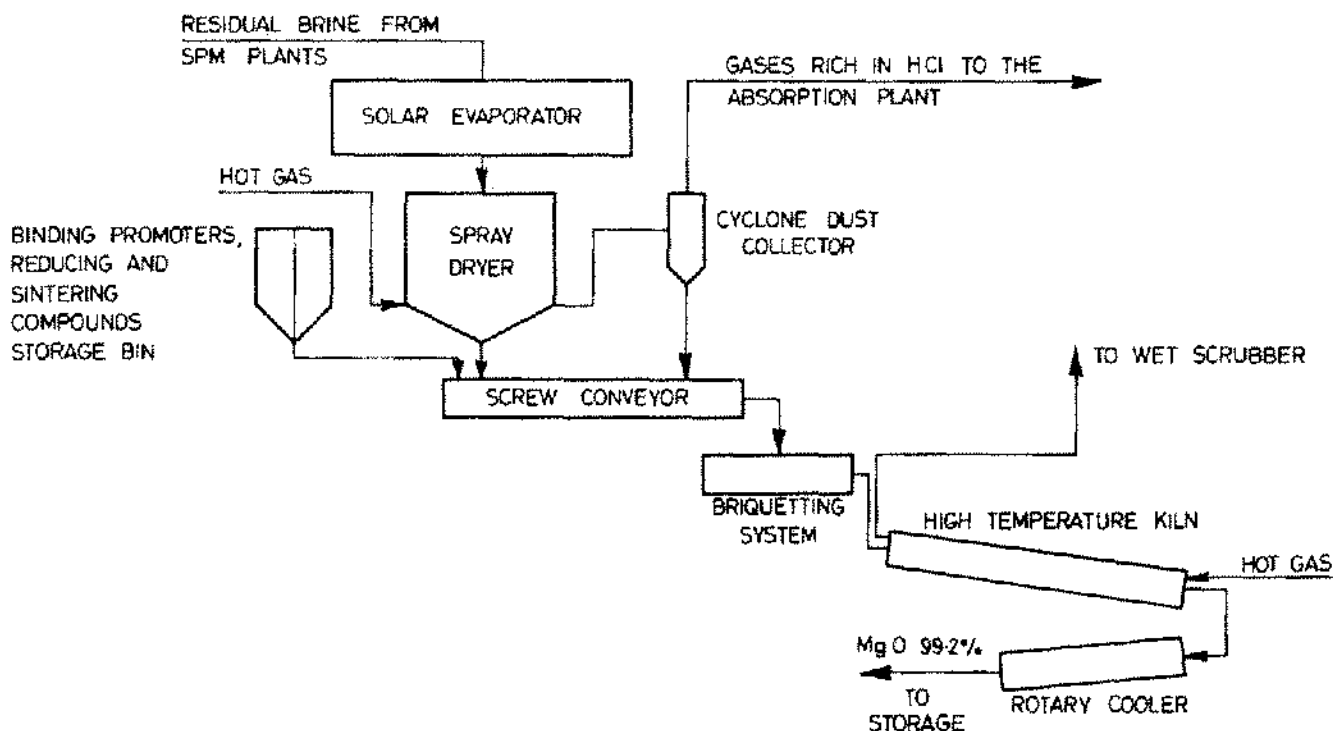


Figure 7. Schematic flow diagram of the proposed process for the production of high purity  $\text{MgO}$  from residual SPM brine.



**Purification of the residual brine.** The residual brines from the SPM and epsomite-carnallite-sylvite plants are returned to the evaporators where they are concentrated to 1.31 specific gravity. In this operation more sodium chloride crystallizes out. The sodium chloride crystals are harvested and the purified brines, with a composition approximate to that shown in Table 1, are pumped to the spray dryer.

**Preparation of the crude magnesium oxide.** In this operation the magnesium chloride contained in the purified brine is converted to  $MgO$ . Decomposition of magnesium chloride hydrates is complete at about 500–550°C, and for this operation a specially designed spray dryer is recommended. The combined decomposition and combustion gases pass through a cyclone dust collector to the hydrochloric acid absorption plant. The product, with a composition approximate to that shown in Table 3, is discharged into a screw conveyor where the binding promoters, reducing and sintering compounds are added and mixed together. The mixture passes into a briquetting system and the hot briquettes are charged into the high temperature kiln.

**Preparation of high purity magnesium oxide.** In this operation the crude magnesium oxide briquettes are calcined at about 1400–1500°C. At this temperature the alkali chlorides volatilize and the magnesium sulphate decomposes, producing a product with a composition similar to that shown in Table 4. The calcination can be accomplished in a continuous rotary kiln. The calciner would be vented to a wet scrubber and exhausted through a stack. The solids are discharged into a rotary water cooler and are conveyed to storage.

**Description of the process for the production of  $Mg(OH)_2$ .** Examining the data presented in this work it is evident that ammonia can be used for the production of high purity  $Mg(OH)_2$  from concentrated seawater brines. Furthermore, the precipitated  $Mg(OH)_2$  has excellent sedimentation and washing properties. These factors offer enough incentive to utilize such results and to incorporate them in a  $Mg(OH)_2$  production process.

The schematic mass flow diagram of the proposed process is shown in Figure 8. The process is divided into six major operations.

**Brine concentration.** In brief this operation comprises the concentration of seawater salt field brine from 1.26 to 1.31 specific gravity in solar evaporation ponds. This step is included in order to concentrate the feed to the ammoniator and, thus, make the process more efficient.

**Ammoniation.** Ammoniation of the concentrated brine can be done in any suitable equipment, such as absorber towers or in special types of crystallizer. Water cooling is necessary to remove the heat of reaction. Crystals of  $Mg(OH)_2$  formed by ammoniation are larger and easier to wash than the crystals obtained by the conventional processes.

**Thicken.** The third step is the thickening of the  $Mg(OH)_2$  brine slurry. This step is included in order to concentrate the feed to the filter and, thus, make the operation more efficient.

**Solids separation.** Solid separation is the fourth step. A hooded rotary filter could be used for the separation. In order to conserve the ammonia and make a high purity product, it is necessary to wash the cake. Seawater and fresh water are used for this purpose. The filter should be vented to an ammonia brine scrubber.

**Drying.** Drying of the product can be accomplished in a continuous, direct-fired, rotary drier such as is common in the chemical industry.

**Ammonia recovery.** The last step to be considered is the recovery of ammonia from the brine. This is accomplished by conventional distillation techniques, which, for the ammonia-water-sodium chloride system, are very well known. Ammonia recovery is of great importance to the economics of the process. The design and operation of the distillation step are essentially the same as that which have been carried out successfully for years in the ammonia-soda process for soda ash production. Therefore, it is likely that the ammonia losses for this process can be controlled within an economical level.

### SOME ADVANTAGES OF THE PROPOSED PROCESSES

It is quite evident that the proposed process for the production  $MgO$  from desulfated seawater brines has several advantages over conventional processes. Some of these advantages are a) simplicity; b) lower fixed and operational costs; c) higher purity of the final product; high magnesium recovery efficiency, and d) production of  $HCl$  as a by-product.

In the proposed process for production of  $Mg(OH)_2$ , the product is obtained on a once-through basis with no brine streams recycled. The feed brine is evaporated to saturation with respect to  $MgCl_2$  and most of the  $NaCl$  is removed by solar concentration. The magnesium recovery efficiency of this process is higher than in the conventional processes. Putting the proposed process on the same production rate basis, the volume of liquid to be processed in the conventional processes is much greater, and the conventional processes demand additional raw materials such as high purity calcined dolomite or limestone, resulting in higher fixed and operational costs. In addition, the proposed process produces larger  $Mg(OH)_2$  crystals, which are easier to wash and of higher purity.

### CONCLUSION

Through direct calcination of purified seawater brines, it is possible to produce high purity  $MgO$ . The results reveal that coal coke is an efficient agent for the removal of im-

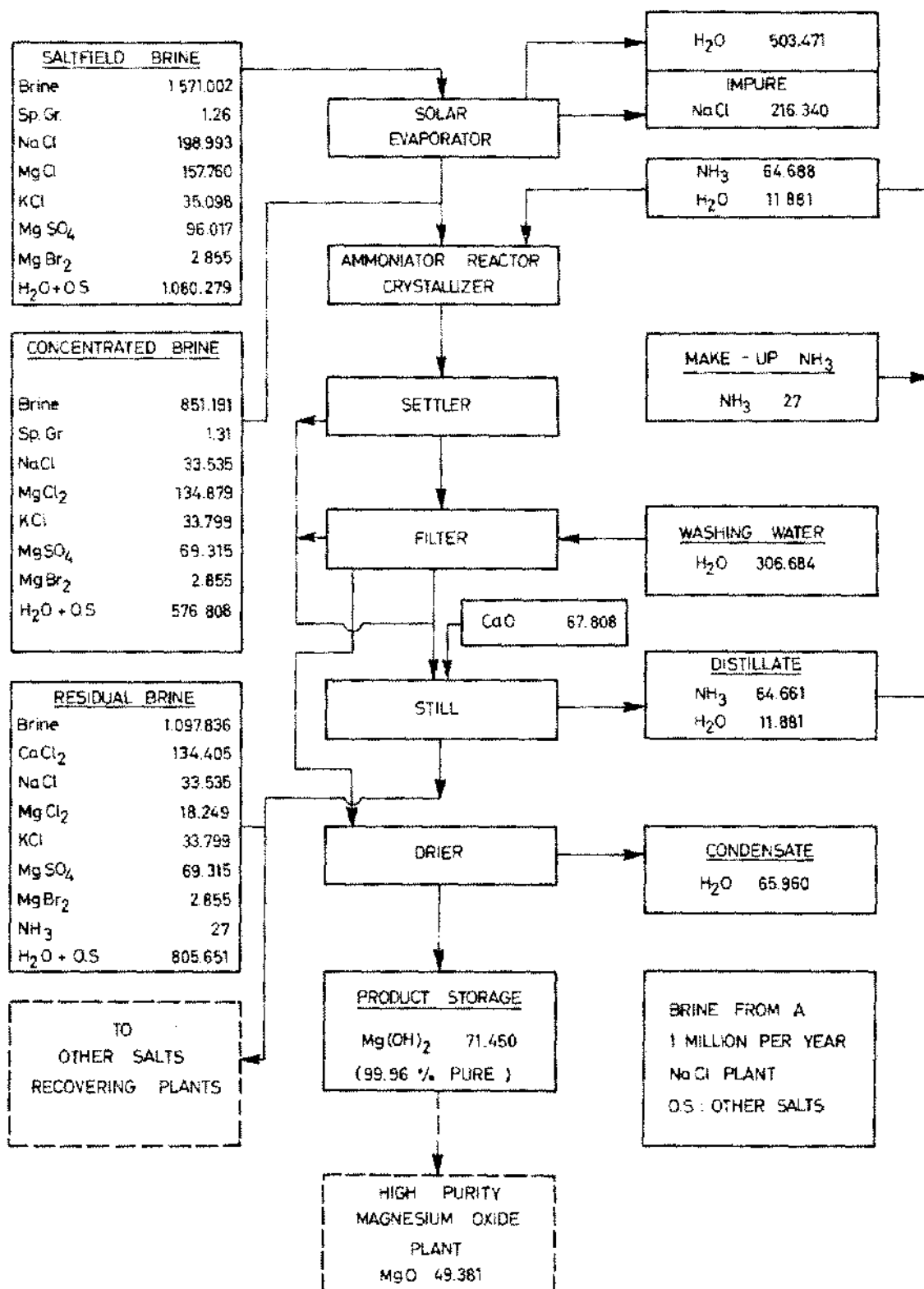


Figure 8. Material balance flow diagram for the production of high purity  $Mg(OH)_2$  from seawater brine by the ammoniation process.

purities such as potassium, sodium, sulfate and boron from crude magnesium oxide during the calcination operation. Higher coke concentration, higher calcination temperature and higher calcination times favor the removal of the impurities and increase the density of the final product.

Employing ammonia as a precipitating agent, it is possible to produce  $Mg(OH)_2$  from concentrated seawater brines to a greater extent than is the case in a conventional process. More important, the  $Mg(OH)_2$  produced is of higher purity. Calcination of this  $Mg(OH)_2$  produces high purity  $MgO$ .

Using the information presented in this work, it is possible to design processes for the production of high purity  $MgO$  and  $Mg(OH)_2$  from concentrated seawater brines. The use of such processes appears to offer important economic advantages and operating flexibilities not possible in the conventional processes.

#### ACKNOWLEDGEMENT

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