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Advanced Coal Power Cycle with a Stand-Alone Magnetohydrodynamic Generator

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A regenerative and recirculation-type coal power system is proposed that works at high efficiencies with a stand-alone magnetohydrodynamic (MHD) generator under complete CO_2 liquefaction. The open-cycle MHD generator is uniquely suited for the proposed cycle because combustion temperatures far exceed the working limits of conventional energy conversion devices due to both high fractional energy recirculation and oxygen combustion required for an efficient CO_2 liquefaction process. The MHD exhaust heat is regenerated sequentially through the following three processes: thermochemical coal gasification, high-temperature syngas preheating, and an iodine–sulphur water splitting process. The mass and energy balances in each elementary process are calculated on the basis of Gibbs' energy minimization principle. It is shown that the attainable system efficiency is much higher than either a coal gasification gas turbine combined cycle or a direct coal-fired MHD combined counterpart under the same evaluation criteria.

Nomenclature

H_i	=	total enthalpy of chemical species i, kJ
h_i	=	molar enthalpy of chemical species i, kJ/kmol
k	=	molar ratio of CO to H_2 in syngas
n_j	=	molar number of chemical element j, kmol
N_i	=	molar number of chemical species i, kmol
Q	=	heat, kJ
T	=	temperature, K
W	=	work, work consumption, kJ
α	=	fraction of magnetohydrodynamic (MHD) exhaust heat split to gasifier
$\Delta H^0_{\rm fs}$	=	standard formation enthalpy of species s, kJ/kmol
η	=	efficiency
θ	=	enthalpy ratio

Subscripts

cb	=	combustor, combustion
cp	=	compressor
gf	=	gasifier
GT	=	gas turbine
IS	=	iodine–sulfur process
LOX	=	oxygen liquefaction, liquid oxygen
M	=	MHD generator
ph	=	preheater, preheated
sep	=	CO_2 separation
0	=	standard state, 0.1 MPa and 298.15 K
(1), (2)	=	inlet, outlet
(ℓ)	=	liquid phase

I. Introduction

AT the present time, the open-cycle magnetohydrodynamic (MHD) generator has the highest available working temperatures of any heat engine device, a clear thermodynamic advantage in terms of high thermal efficiency capability. On the other hand, the alkali metal seeded MHD generator must operate with combustion plasma temperatures above 2400 K to retain an effective level of the electrical conductivity, and the high-enthalpy exhaust is a significant adverse feature. It is, therefore, a natural consequence that the remaining enthalpy be regenerated by a steam turbine cycle and that the required plasma temperature be achieved through combustion with high-temperature air that has been preheated by the MHD exhaust. Note that the MHD topping and steam turbine bottoming combined cycle was a major target of commercial systems in the past research and development activities for both gas-fired and coal-fired MHD power generation.

In clean fuel-fired MHD combined systems, air preheating can be performed by a regenerative heat exchanger consisting of heat accumulation-type ceramic piles in which air is heated directly in contact with the heater elements up to about 2000 K. The preheated oxidizer enthalpy, together with the fuel heat value, mainly determine the temperature and electrical conductivity of the working plasmas. In the case of natural gas–air combustion plasmas seeded with K_2CO_3, the temperature and the conductivity are at most 3000 K and 10–15 S/m, respectively. Under such working conditions, the system efficiency was estimated to be around 60% (Ref. 1) without CO_2 recovery. It was considered that the efficiency upgrade was the major advantage of gas-fired MHD power generation for reduced cost of electricity, fuel saving, and environmental countermeasure. Natural gas-fired MHD research in the former U.S.S.R. resulted in the planning and subsystem construction of U-500, the first commercial scale MHD–steam combined plant.[2] The plant has not been completed, however, because of the drastic change in social and economical situations. From the technological point of view, note that the serious problem of low conductivity, even in a gas-fired plasma, resulted in a massive design effort to increase efficiency. The U-500 final design, for instance, required a channel length of 20 m with a super-conducting magnet weighing 5000 tons (Ref. 3).

On the other hand, the coal-fired MHD combined cycle was the main direction of research and development (R&D) activities in several countries from 1988 to 2000. Among them, the U.S. Department of Energy (DOE) carried out a world leading research and development program for coal-fired MHD retrofit of existing coal power plants during 1984–1993. The major objectives of DOE's proof-of-concept (POC) program were to reduce the cost of electricity generation and to reduce SO_2 emission for the environmental benefit. Almost of the program goals were met for integrated MHD topping subsystems using 50-$MW_{\rm thermal}$ coal combustor at the Component Development and Integrated Facility in Butte, Montana.[4] The bottoming cycle counterparts were demonstrated at the 28-$MW_{\rm thermal}$

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Coal Fired and Flow Facility in Tullahoma, Tennessee.[5] Unfortunately, technological and economical risks from an electric utility's viewpoint remained unanswered.

Problems such as the incompatibility of a ceramic pile-type air heater with slag, highly stressed anode and cathode walls, and low electrical conductivity, for example, seem difficult to solve in direct coal-fired schematics because their origin lies in the behaviors of coal slag and slag vapor in topping components. The ionic current transport in slag layers on electrode walls severely induces anode oxidation and material wear, and the cathode shorting leads to decreased output performance.[6] The low air preheat temperature provided by a radiant-type heat exchanger (approximately 1000 K) requires additional oxygen-enrichment to achieve an appropriate plasma conductivity. Even with oxygen-enriched combustion, the coal-fired plasma temperature was generally lower than the gas-fired counterpart because of large continuum radiation loss from slag particulates. Such insufficient conductivity levels may require a sizable channel and superconducting magnet. The plasma performance value defined by the power density divided by the magnetic flux density squared in the POC 1A4 generator design,[7] for instance, was approximately 0.69 $MW_e/m^3/T^2$. This was less than one-half of the typical value for clean fuel-fired plasma based on a conductivity of 10 S/m, a velocity of 1000 m/s and a load factor of 0.8.

The design of a base-load-type, open-cycle MHD power generator should, therefore, be directed at slag-free operation with elevated combustion temperatures and electrical conductivity in comparison to typical values for gas-fired plasmas. With respect to the fuel, coal should still be the desired option for future fossil electrical power generation because of its abundance. Thus, both efficiency upgrade and CO_2 reduction will be of major concern to future MHD central power generation systems.

The authors have recently proposed a stand-alone scheme for an open-cycle MHD system incorporating thorough regeneration and energy recirculation features,[8] where the mass and energy balances in the combustion and thermochemical regeneration processes were evaluated on the basis of overall chemical reactions. Moreover, a steam generation unit was introduced between the MHD generator and the downstream gasifier to regulate the combustion temperature in the topping combustor.

This analysis is extended in the present work, where the steam generation branch need not necessarily be considered. Here, the efficiency is estimated with a more rigorous model of equilibrium gas components, and the thermodynamical properties in the combustion, gasification, and high-temperature syngas preheating processes are calculated on the basis of the Gibbs potential minimization principle.

II. Design Concept for an Efficient Thermal Cycle

Steam generation and combustion air preheat comprised the basic regeneration concepts of past MHD power generation systems for upgrading efficiency in both coal-fired and gas-fired schema. In combined systems, however, a definite amount of regenerated steam heat is lost in the bottoming steam cycle. Therefore, the efficiency of the combined cycle has an inherent limit.

In principle, the attainable efficiency limit can be eliminated in stand-alone schematics with thorough heat regeneration and energy recirculation. The energy flow schema in the regeneration and recirculation-type cycle and in the combined counterpart are schematically shown in Figs. 1a and 1b, respectively, and each efficiency characteristic is shown in Fig. 1c. Here, the regeneration factor R is defined as the ratio of heat recirculated (regenerative design) or the steam heat (combined design) to the heat input

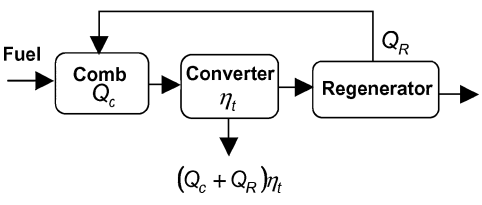

Fig. 1a Energy flow in regenerative scheme.

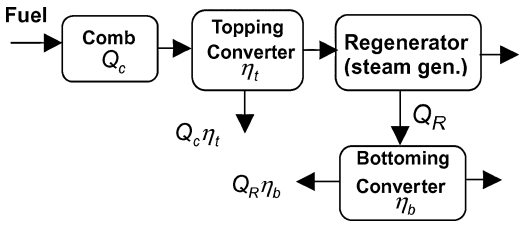

Fig. 1b Energy flow in combined scheme.

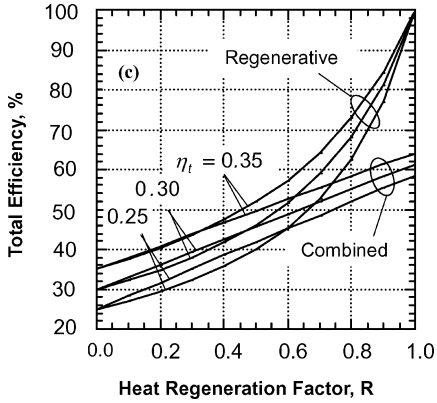

Fig. 1c Total efficiency dependence on the heat regeneration factor.

to the regenerator, as given by $R = Q_R/[(Q_c + Q_R)(1 - \eta_t)]$ and $R = Q_R/[Q_c(1 - \eta_t)]$, respectively. As we can see in Fig. 1c, with the same topping converter efficiency η_t and the fuel heat value Q_c, the total efficiency defined by the ratio of work output to the fuel heat value, is much higher in the regenerative system than in the combined system under high regeneration factors. It is also shown that the regenerative system efficiency is less dependent on η_t under high R values. It is clear from the energy conservation law that the combustion temperature in the regenerative system is much higher than in the combined system. Therefore, an open-cycle MHD generator might be the only possible conversion unit in the regenerative cycle because it involves no mechanically rotating parts. Furthermore, the high-temperature combustion characteristics may solve substantial problems encountered in past MHD technology development programs.

III. Stand-Alone Scheme of MHD Power Generation System

Here, assuming coal as the primary fuel, we consider three sequential processes for the regeneration of MHD exhaust heat of wide temperature range. The first process adopted for the temperature range above 1700 K is thermochemical gasification, as described by

$$C + CO_2 = 2CO + 2\Delta H^0_{fCO} - \Delta H^0_{fCO_2} \quad (1)$$

$$C + H_2O = CO + H_2 + \Delta H^0_{fCO} - \Delta H^0_{fH_2O} \quad (2)$$

Here the heat of the reactions is supplied from the exhaust of syngas–oxygen combustion products coming from the MHD unit. Unlike the conventional gasification, no partial combustion of coal is considered, and, therefore, the reaction heat absorbed as the syngas chemical energy amplifies the combustion heat of the primary coal by roughly the amount of heat absorbed. In the case of the reactions (1) and (2), the syngas combustion heat is about 1.35 times higher than that of coal. This is one origin of efficiency upgrade in the regenerative, stand-alone system.

Assuming that all slag can be rejected at the gasifier exit under the ash melting temperature, we can adopt a heat accumulation-type second regenerator, where the heat of the slag-free syngas/CO_2/H_2O mixture in a temperature range of approximately 1300–2400 K can be regenerated as the latent heat of syngas fuel consisting of CO

and H_2 coming from the CO_2 separator. The potassium seed is in the vapor phase within this temperature range, so that, from the working temperature point of view at least, the adoption of ceramic-pile-type heat exchangers may be tolerable based on extensive experience in the U-25 facility.[9]

The third process, useful below 1300 K, is the iodine–sulfur process (IS-process), proposed by Norman et al.[10] The IS-process is a water splitting reaction comprising three separate reactions described by: $I_2 + SO_2 + 2H_2O = H_2SO_4 + 2HI$, $H_2SO_4 = H_2O + SO_2 + \frac{1}{2}O_2$, and $2HI = H_2 + I_2$. The overall reaction can be considered as a water decomposition process, $H_2O = H_2 + \frac{1}{2}O_2$, with heat absorption of 13.8×10^3 kJ/kmol. The advantage of the IS-process is that each process can be operated in gaseous and/or liquid phase, the sulfur and iodine chemicals circulate in closed loops, and no pollutants are exhausted into the atmosphere. Because this process is currently under development for hydrogen production purposes by the use of heat from a high temperature gas-cooled nuclear reactor at the Japan Atomic Energy Research Institute,[11] it might be useful for the large amount of heat to chemical energy regeneration required in a large scale, base-load-type system. The IS-process efficiency defined by the ratio of the hydrogen combustion heat to the input heat is currently estimated to be 40–50% (Ref. 12). There is also an upgraded efficiency capability by modification of the O_2 separation process.[13]

The proposed stand-alone MHD power generation system is shown in Fig. 2, where the primary coal is synthesized thermochemically with combustion products and input water according to the reactions (1) and (2) in a gasifier installed downstream of the MHD diffuser. The combustion takes place with recycled syngas and oxygen, so that no slag is involved in the MHD flow train. Therefore, all problems that were associated with coal slag in the direct coal-fired MHD system can be removed in the present system.

An appropriate fraction $(1 - \alpha)$ of the MHD exhaust should bypass the gasifier and be mixed with the syngas/CO_2/H_2O mixture emerging from the gasifier before entering the preheater. Note that no slag is involved in the mixing process because the bypassed gas is slag free. Because of the definite temperature drop associated with the gasification process, the heat split is necessary to match the gasifer exit temperature to the slag removal temperature, which is assumed to be 1700 K. On mixing the exit gas with the bypassed combustion products, the mixture temperature rises far beyond the values that might be compatible with state-of-the-art heat exchanger materials. Thus, we assume that the temperature can be fixed at 2400 K at the point ***N*** by introducing low-temperature gas from the exit of the IS-process. We note that none of the new chemical element is introduced by mixing the cooling gas. The gaseous heat is regenerated as latent heat of the CO_2-free syngas within the heat exchanger and further regenerated as the hydrogen chemical energy within the IS-process. The temperature at the IS-process inlet is also fixed at 1300 K at point ***M*** by the introduction of an appropriate fraction of the IS-exhausts. After exiting the IS-system, H_2O and K_2CO_3 particulates are separated in the filter stage. To minimize the ambiguity in energy balance calculations, and also for comparative discussions with other systems, no seed regeneration/reforming processes are considered in the present system. Note, however, that the seed recovery and reprocessing may be much easier and more economical than in the direct coal combustion MHD system because slag and seed can be extracted from separate units in principle on the basis of the phase-change temperature difference in the present system.

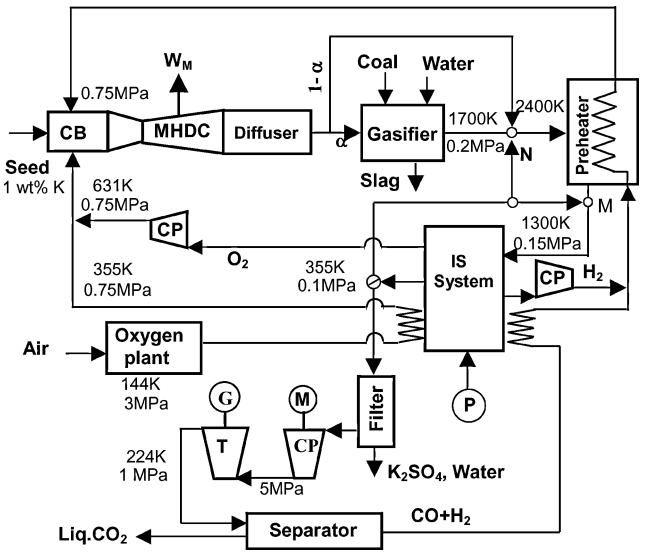

Fig. 2 MHD stand-alone system: CB combustor, MHDC MHD channel, CP compressor, T turbine, G turbogenerator, M motor, and P pump.

After the filter stage, syngas is separated from CO_2 in a separation unit comprising an isothermal compression and an adiabatic expansion. This dual process can be applied to the syngas/CO_2 mixture on the basis of phase-change temperature difference in CO, H_2, and CO_2. First, we assume that the mixture is compressed from 355 K and 0.1 MPa to 355 K and 5 MPa and then expands in a turbine to a 224-K and 1.0-MPa state. The carbon dioxide can then be liquefied under the temperature and pressure at the turbine exit, while CO and H_2 remain in the gaseous state, so that some fraction of compression work can be recovered. After exiting the CO_2 separation unit, the CO_2-free syngas is preheated by IS-exhaust heat to 355 K, mixed with the IS–hydrogen product, and then heated further in the regenerative preheater. Finally, the preheated syngas is recirculated to the combustor where the combustion should be performed with pure oxygen. The IS–oxygen could also be used for combustion. Oxygen combustion should be considered primarily for an efficient CO_2 separation. Moreover, it has essential effects on the MHD conversion processes, namely, elevation of the combustion temperature and electrical conductivity, high power density, reduced mass flow rate, and reduced channel size and magnet volume. The combustor pressure is fixed at 0.75 MPa in the present design, and the theoretical enthalpy extraction ranges up to 31.2–34.4%, depending on the combustion temperatures when the channel exit pressure is fixed at 0.2 MPa. We suppose that the MHD channel enthalpy extraction ratio can be varied arbitrarily up to 30% under fixed combustor pressure.

In actual system designs, several modifications should be taken into account, especially for regeneration of cooling heat from the high-temperature components. A possible regeneration scheme may be the heating of feed water introduced into the gasifier and/or the utilization of liquid oxygen vaporization. Because all regenerated heat contributes to an increase in the working gas enthalpy in the stand-alone scheme, the modification may result in about the same system performance as estimated under the adiabatic wall assumption. We neglect all heat losses in the present paper because the major objective is to demonstrate a new capability of the open-cycle MHD power generation on the basis of comparative discussions with past studies on conventional combined schema and/or on an MHD–steam combined system designed with the thermochemical synthesis and with the same evaluation criteria.[14]

IV. Estimation of Efficiency

In view of the major objectives of the present paper, we introduce the following assumptions: 1) All gaseous compositions and properties upstream of the IS-process can be estimated by chemical equilibrium analysis by the use of the element-potential method based on Gibb's energy minimization principle. 2) All heat losses are neglected and the energy balance in each unit can be estimated by the inlet and outlet enthalpy differences. 3) The gaseous compositions are in a frozen state below 1300 K. 4) Heat losses accompanying slag removal are neglected. 5) Temperature and pressure values fixed in the calculation are those as indicated in Fig. 2.

First, we note that in each reaction of carbonaceous fuel reforming with carbon dioxide and steam, reactions (1) and (2), or a reaction under lower temperature described by $C + 2H_2O = CO_2 + 2H_2$,

the sum of molar numbers of CO and H_2 is twice the carbon molar number in the primary coal, even if the shift reaction $CO + H_2O = CO_2 + H_2$ is considered. Therefore, the sum of CO and H_2 obtained from the gasification of 1 kmol carbon input is given by $N_{CO} + N_{H_2}^{GF} = 2$. Assuming that the molar ratio of CO to H_2 is given by k, we obtain $N_{CO} = 2k/(k+1)$ and $N_{H_2}^{GF} = 2/(k+1)$. According to chemical equilibrium calculation, k is approximately equal to two above 1500 K.

The adiabatic flame temperature T_{CB} can be determined from the conservation of total enthalpy: $\sum H_i = N_{CO} h_{CO}^{PH(2)} + N_{H2} h_{H_2}^{PH(2)} + N_{O2} h_{O_2}^{CB(1)} + N_{seed} h_{seed}^0$, where $\sum H_i$ is the total enthalpy of the combustion products given by the equilibrium calculation. The right-hand side represents the total enthalpy of the reactants. The molar numbers are given by $N_{H_2} = N_{H_2}^{GF} + N_{H_2}^{IS}$ and $N_{O_2} = (N_{CO} + N_{H_2})/2$. The molar number of K_2CO_3 seed material is estimated under the assumption of one mass percent of K in the total mass input. The combustion heat is calculated from the total enthalpy of the combustion product relative to the standard state as $Q_{CB} = \sum H_i^{CB(2)} - H^0$. Here, $H^0 = N_{H_2} \Delta H_{fH_2O}^0 + N_{CO} \Delta H_{fCO_2}^0$. Then, the total net electrical work of the system can be estimated as $W = Q_{CB}\eta_M - W_{LOX} - W_{sep} - W_{CP}$, where the first term is the electrical work output from the MHD channel. The second to fourth terms represent the work consumption associated with liquid O_2 production supplied from the liquid oxygen storage, CO_2 separation, and compression of IS hydrogen and oxygen, respectively, where W_{LOX} is assumed to be 9.89×10^3 kJ/kmol, referring to the present commercial value. We assume the CO_2 separation turbine exit pressure is sufficient to recirculate the syngas to the combustor via the two preheating processes. With the preceding assumptions and calculations, the total system efficiency is estimated by $\eta = W/(-\Delta H_{fCO_2}^0)$.

In the gasifier, the energy balance can be written as $\alpha\{(1-\eta_M)Q_{CB} + H^0\} + H_C^0 + H_{H_2O(\ell)}^0 = \sum H_i^{GF(2)}$. The left-hand side is the total input energy and the right-hand side is the gasifier exhaust energy estimated for the given temperature and pressure values of 1700 K and 0.2 MPa, respectively. Note that the absorbed reaction heat is evaluated as the syngas chemical energy in the right-hand side enthalpy. Also, we note that a mass balance condition $\alpha(N_{CO} + N_{H_2}) + N_{H_2O(\ell)}^{GF} \geq N_C$ should be met for completion of gasification, where $N_{H_2O(\ell)}^{GF}$ is the number of water species to be introduced into the gasifier when α is small, that is, when η_M is low.

After the gasifier, the syngas products are mixed with the bypassed MHD exhaust. On mixing, the temperature increases and the equilibrium composition changes. Because the temperature exceeds the allowable value of the regenerative-type heat exchanger, we fix the temperature as 2400 K at the preheater inlet. This is accomplished by the introduction of a part of the low-temperature IS-process exhaust. Here, for the purpose of further computation, we denote the mixture before mixing with the IS-exhaust as the basic system and its molar number of the chemical element as n_j (j = C, H, O). The temperature of the basic system depends on the MHD channel enthalpy extraction η_M and the fraction of the heat split α. Although the top temperature of the IS-process may not necessarily be constrained in principle, we fix the inlet temperature at 1300 K by also mixing an appropriate IS-exhaust gas, taking into account the state-of-the-art IS-process technology.

The total enthalpies of the basic system include the preheater inlet $H_{basic}^{PH(1)}$, the IS-process inlet $H_{basic}^{IS(1)}$, and the exit $H_{basic}^{IS(2)}$ and are calculated at the specified temperature and pressure. The enthalpy H_{basic}^0 at the standard state can be calculated with molar numbers given by $N_{CO_2}^0 = 1$, $N_{CO}^0 = 2k/(k+1)$, $N_{H_2}^0 = 2/(k+1)$ and $N_{H_2O(g)}^0 = N_{H_2}^{IS} + N_{H_2O(\ell)}^{GF}$. Then, the energy and mass balance at the preheater inlet can be written as $\sum H_i^{PH(1)} = H^{GF(2)} + (1-\alpha)H^{M(2)} + \theta_1 H_{basic}^{IS(2)}$ and $n_j^{PH(1)} = (1+\theta_1)n_j$, where θ_1 is the ratio of enthalpy of the cooling gas to that of the basic system. Because the fraction of the chemical element n_j is the same in both the basic system and the cooling gas, the enthalpy of the introduced cooling gas at the preheater inlet state is given by $\theta_1 H_{basic}^{PH(1)}$. Thus, the enthalpy at the preheater inlet can be rewritten as $\sum H_i^{PH(1)} = (1+\theta_1)H_{basic}^{PH(1)}$. In the preheater, the syngas fuel is heated up to T_{PH}. The absorbed heat of fuel is estimated as $Q_{PH} = \sum H_i^{PH(1)} - \sum H_i^{PH(2)}$. Denoting the ratio of enthalpies of the cooling gas to that of the basic system at the point M as θ_2, we obtain the energy and mass balance at the inlet of the IS-process as $\sum H_i^{PH(2)} + \theta_2 H_{basic}^{IS(2)} = (1+\theta_1+\theta_2)H_{basic}^{IS(1)}$ and $n_j^{IS(1)} = (1+\theta_1+\theta_2)n_j$. Then, the heat input into the IS-process is estimated by $Q_{IS} = \sum H_i^{IS(1)} - (1+\theta_1+\theta_2)H_{basic}^0$, and, therefore, the molar number of the hydrogen from IS-process is given by $N_{H_2}^{IS} = Q_{IS}\eta_{IS}/(-\Delta H_{fH_2O}^0)$ for given IS-process efficiency. Using the energy and mass balances thus far derived at the mixing points N and M, we can calculate T_{PH} and $N_{H_2}^{IS}$ iteratively.

V. Results and Discussion

The calculated system efficiency is shown in Fig. 3 as a function of the MHD channel enthalpy extraction and the IS-process efficiency. Also indicated in Fig. 3 are the efficiencies of a combined MHD–steam system incorporating thermochemical synthesis and regenerative syngas preheater concepts, but without an IS-process. The performance of the MHD combined system was originally estimated on the basis of overall reactions in the gasification and combustion processes in Ref. 14. The values shown in Fig. 3 are those reevaluated by estimation of the efficiency drops between the overall reaction model[8] and the equilibrium counterpart[15] as applied to the same stand-alone MHD scheme. In the combined case, the effect of CO_2 recovery was estimated by a 4% point drop in the total efficiency by the use of the CO_2 separation model employed in Ref. 14 and under the assumption of a 48.5% steam turbine efficiency, as defined by the ratio of steam heat to turbine work. The same conditions were employed for the combustor pressure, with a disregard for heat losses and the lowest temperature and pressure in the system. We note that the thermochemical combined scheme may be favorable when the MHD channel enthalpy extraction is low. On the other hand, much higher efficiencies can be obtained in the stand-alone system when both the enthalpy extraction and IS-process efficiency are sufficiently high. Presumably, this is achievable with current state-of-the-art technologies. The efficiency dependence on the heat regeneration factor can be explained by the characteristics of the regenerative scheme discussed in Sec. II. This can also be clearly understood as a consequence of the high heat regeneration factor in the present system as shown in Fig. 4. The maximum system efficiency obtained in the present system for $\eta_M = 0.3$ and $\eta_{IS} = 0.5$ is about $\eta = 0.59$ under full CO_2 liquefaction. Although the assumed MHD enthalpy extraction may be the maximum attainable value from a practical perspective, further gains in system efficiency are available with additional increases in IS-process efficiency.

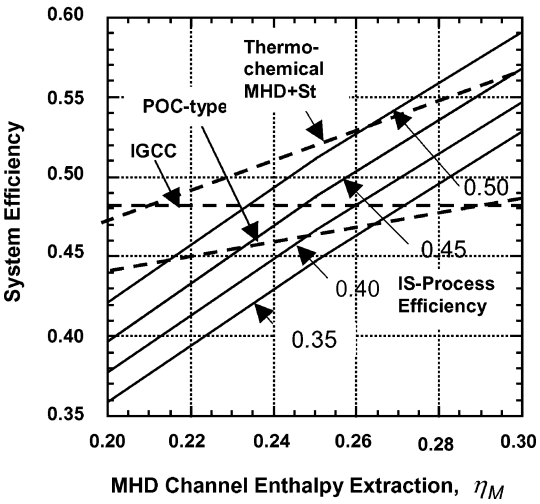

Fig. 3 Efficiency characteristics of stand-alone MHD system as function of the MHD channel enthalpy extraction and the IS-process efficiency.

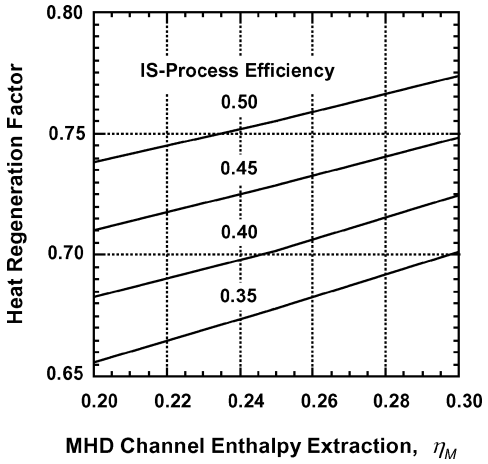

Fig. 4 Ratio of heat regenerated to the MHD exhaust heat as function of channel enthalpy extraction and IS-process efficiency.

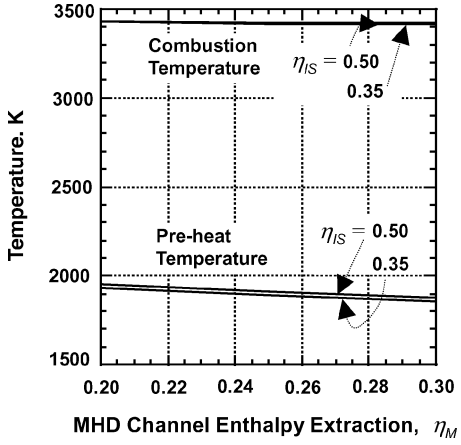

Fig. 5 Combustion temperature and fuel preheat temperature as function of the MHD enthalpy extraction and IS-process efficiency.

Comparing efficiency attributes with the integrated gasification combined cycle (IGCC) and POC-type MHD–steam combined systems estimated using overall reactions in elementary processes,[14] we can say that the efficiency of the IGCC system is lower by about 6% and that of the POC-type by 4.5% when the same enthalpy extraction ratio is employed for the latter. The CO_2 liquefaction is not considered in these cases. If its effect were considered and performance were estimated on the basis of chemical equilibrium compositions and thermodynamic properties as done in the present calculation, the efficiency of the IGCC- and POC-types would fall below that of the present system by more than 10%.

A distinctive feature of the CO_2-free, stand-alone MHD system is its extremely high combustion temperatures (Fig. 5). The temperature ranges above 3400 K and is almost independent of the MHD channel enthalpy extraction and the IS-efficiency. This derives from the temperature constraints imposed on the fuel preheater and IS-process inlet points. The results suggest that the MHD stand-alone scheme has the potential for wide range selectivity in combustion temperature by changing the design values at appropriate points. If the temperature constraints are not imposed, the combustion temperature becomes strongly dependent on the channel enthalpy extraction and the heat split value, as discussed in Ref. 8. The high-temperature feature of the present system is attributed to oxygen combustion, efficient chemical and thermal energy regeneration, a high fuel preheat temperature (Fig. 5), and energy recirculation to the combustor. When we define the heat amplification factor as the ratio of total channel heat input Q_{CB} to that of the direct carbon combustion ($-\Delta H^0_{f\mathrm{CO}_2}$), its value ranges from 2.3 to 2.5 with $\eta_{IS} = 40$–50% and $\eta_M = 25\%$.

VI. Conclusions

The advantage of a regenerative and recirculation-type MHD energy system in regard to thermal efficiency upgrade and CO_2 emission was numerically demonstrated by the use of the heat regeneration factor as a key parameter. The technological achievements and unsolved problems of past gas-fired and direct coal-fired MHD power generation activities were reviewed, and it was pointed out that the critical problems were low plasma conductivity in the gas-fired systems and low conductivity plus coal slag in the coal-fired system topping components. It was emphasized that the problems could be solved, in principle, in a fully regenerative scheme, and that a syngas-fired open-cycle MHD generator has a high potential for clean coal utilization when it is adopted for the power generation unit in the proposed system concept. At present, the open-cycle MHD generator appears to be the best option for the exploitation of the thermodynamic advantage of high combustion temperature in the stand-alone, regenerative cycle.

In the stand-alone MHD system of this paper, three stage heat regeneration processes were considered, and MHD exhaust heat of approximately 70% was regenerated and recirculated to the combustor as chemical energy and latent heat of the coal-synthesized fuel. On the basis of energy conservation, the combustion temperature with oxygen was in the range of 3450–3500 K for the assumed numerical model. The thermochemical coal synthesis and high working temperature suggested a new advanced MHD power generation capability where the inherent physical and technological problems associated with coal slag in the direct coal-fired schematics could basically be eliminated, and the low conductivity feature encountered in the gas-fired plasmas could be greatly improved. It is certain that the thermodynamic advantage of an open-cycle MHD generator could be achieved by the use of a regenerative and recirculation scheme in a stand-alone system.

The system efficiency obtained was on the order of 60%, even when a complete CO_2 liquefaction was taken into account. This value is far beyond current performance targets for integrated coal gasification combined cycles and both gas-fired and coal-fired MHD–steam combined systems where a low-power CO_2 liquefaction process cannot be implemented because of complexity in gaseous composition related to air combustion. Furthermore, economical CO_2 recovery could be achieved with a coal-powered system with a stand-alone MHD generator unit driven by syngas–oxygen combustion products.

Many technological challenges need to be cleared to obtain this goal. However, note that several key issues may never be far beyond the scope of state-of-the art technologies. The database and experiences accumulated in current IGCC activities and those in past MHD research activities, such as high-temperature air preheater technology, high-interaction MHD channels, and high-temperature materials developed for hot electrode and insulator wall designs, may be especially useful in the future development of an advanced MHD system for an economical and environmentally acceptable coal power cycle.

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