TECHNO-ECONOMIC MODELLING OF A UTILITY-SCALE REDOX FLOW BATTERY SYSTEM

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ABSTRACT

A one-dimensional numerical model has been developed for redox flow battery (RFB) systems with bipolar flow-by electrodes, soluble redox couples, and recirculating batch operation. Overpotential losses were estimated from the Butler-Volmer equation, accounting for mass-transfer. The model predicted the variation in concentration and current along the electrode and determined the charge-discharge efficiency, energy density, and power density. The model was validated using data obtained from a pilot-scale polysulphide-bromine (PSB) system commercialized by Regenesys Technologies (UK) Ltd. The model was able to predict cell performance, species concentration, current distribution, and electrolyte deterioration for the Regenesys system. Based on 2006 prices, the system was predicted to make a net loss of 0.45 p kWh^{-1} at an optimum current density of 500 A m^{-2} and an energy efficiency of 64%. The economic viability was found to be strongly sensitive to the kinetics, capital costs, and the electrical energy price differential.

Keywords: Redox flow battery, polysulphide bromine, techno-economic modelling

INTRODUCTION

Redox flow batteries (RFBs) have been investigated for many years as chemical stores of electrical energy [1], and are the closest storage technology to widespread commercialization. RFBs have numerous advantages over other batteries, including a separation of the energy and power rating, modular systems, repeatable cyclic behavior, and the use of benign chemicals at atmospheric temperature and pressure. Redox couples currently under development for use in RFBs include polysulphide-bromine (PSB) [2], vanadium-vanadium [3], vanadium-polyhalide, cerium-zinc [4] and lead-lead. Numerical modelling of RFB systems for energy storage applications allows the technical and commercial performance of different designs to be predicted without costly lab, pilot, and full-scale testing.

In this paper we develop a numerical model of a RFB system, and apply this to the Regenesys Technologies Ltd. Pilot-scale PSB-based RFB. Numerical modelling can be used to obtain key parameters such as the electrochemical rate constants for the reactions. Furthermore, once validated, the model has been used to evaluate and optimize the design and performance of a full-scale commercial RFB system.

METHODOLOGY

The RFB numerical model developed in this paper performed the following functions:

- Evaluation of mass transport (from the electrolyte bulk to the electrode surface) and reaction kinetics (described by the Butler-Volmer equation) as rate-determining processes.
- Estimation of the variation in concentration, overpotential, current density, exchange current density, and limiting current density up the electrode in a one-dimensional model.
- Calculation of the variation in cell performance during charge-discharge cycles and overall system characteristics including energy efficiency, power density, and energy density.
- Consideration of different operating conditions and electrolyte systems, e.g., variable redox couple, applied current density, power rating, operating temperature, catalyst, cycle length, species

concentration, electrolyte velocity, electrode area, stack size, tank volume, electrolyte conductivity, and membrane conductivity.

The main assumptions used in the model are listed below:

- Single-step electrochemical reactions involving dissolved electro-active species were assumed to occur at the electrodes.
- Electrochemical kinetics were assumed to be described by the Butler-Volmer equation.
- Electrochemical rate constants and mass-transfer coefficients were assumed to be approximately constant.
- The effects of electro-migration were assumed to be negligible.
- Adsorption, electrode resistivity, and shunt current effects were assumed to be negligible.
- The current efficiency was assumed to be 100%, i.e., side reactions were not considered.
- Plug flow conditions were assumed to occur in the cells.
- Conditions were assumed to be the same in each cell in a stack.

The energy storage plant specification was based on the first utility-scale PSB storage plant constructed by Regenesys Ltd. (although for commercial reasons the plant was never commissioned). The plant was constructed using XL200 modules, which consisted of 200 XL cells assembled to form a bipolar stack. These modules were rated at 120 kilowatts (kW) assuming a current of 400 amperes and a cell voltage of 1.5 volts. The plant was specified to give a power output of 15 megawatts (MW) and an energy storage capacity of 120 megawatt hours (MWh) (corresponding to an 8-hour discharge). Sodium bromide (NaBr) and $Na_2S_{4.8}$ electrolytes were used in the sulphide and bromide tanks respectively, with $[Br]_{T,BOC} = 4.5$ M and $[S]_{T,BOC} = 4.8$ M. The electrolyte volume was adjusted in order to deliver 120 MWh of capacity.

In terms of lifetime, the limiting component was expected to be the membrane, which typically lasts 15 years in the harsher chlor-alkali industry. The plant life was therefore assumed to be 15 years, with around 250 cycles of utilization per year. While the model was capable of modelling self-discharge over a long series of cycles, the effect of self-discharge was not included in this study. In practice, electrolyte conditioning would be required to keep the system in balance. It was therefore assumed that conditioning was carried out regularly, so that the same performance could be expected on each cycle.

Obtaining accurate data on the capital cost of process equipment, particularly for a technology that has not been established, is extremely challenging. For this study, an approximate capital cost has been estimated based on a "best-case" scenario for the commercial performance, assuming a mature RFB industry has been established. The capital cost model was based on an existing model of a PSB system that predicted a capital cost of £320 kW^{-1} [5]. This model assumed 1995 UK prices, a production rate of 400 MW per year of 200 kW-rated modules, mature production costs (representing the middle of the growth phase of the product life cycle), and modularization and standardization of plant designs. It was assumed that increased costs due to inflation since 1995 would be offset by savings associated with technological improvements. The capital cost is divided into three elements: electrochemical cells (including balance of plant), electrical equipment and electrolyte and tanks [5]. The economic parameters used are shown in Table 1.

Table 1. Parameters used in the economic model.

Overall	Required delivered energy	120 MWh per cycle at 15 MW
	Charge/discharge period	8 hours
	Frequency of charge cycles	250 cycles $year^{-1}$
	Plant lifetime	15 years
Capital cost	Installed module capital cost	£41,000 $N_m^{0.9}$
	Electrical plant cost	£60 kW^{-1} on charge
	Electrolyte/tank cost	£350 m^{-3}
Running Costs and Income	Pump efficiency, π_p	35%
	Transmission losses	5 % on discharge
	Cost of electricity consumed	2.3 p $kWhr^{-1}$
	Value of electricity delivered	5.7 p $kWhr^{-1}$
Net Present Value Calculation	Inflation rate, *a*	2.5%
	Discount rate, *r*	10%
	Net present value factor	9.25 years

RESULTS AND DISCUSSION

The overall cost of the delivered energy was calculated for a range of operating current densities (Figure 1). At low current density the cost of the electrochemical modules was the largest part of the cost. As the current density increased, this cost decreased rapidly, but at high current density the cost of inefficiency (i.e., energy lost on charging) increased significantly. Thus an optimum current density of around 500 A m^{-2} was obtained.

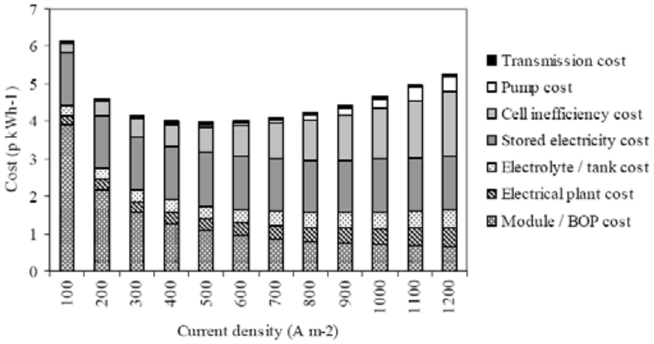

Fig. 1. Breakdown of the cost of delivered energy at a range of current density.

For arbitrage applications, the electricity costs shown in Table 1 were estimated based on UK electricity market prices from 2006. Figure 2 shows the net profit obtained showing that the process is uneconomic for the conditions studied. However, the loss is less than 0.5 p kWh^{-1} at the optimum current density and it is likely that improved performance and/or changes in the electricity market are likely to make the technology profitable within the next decade

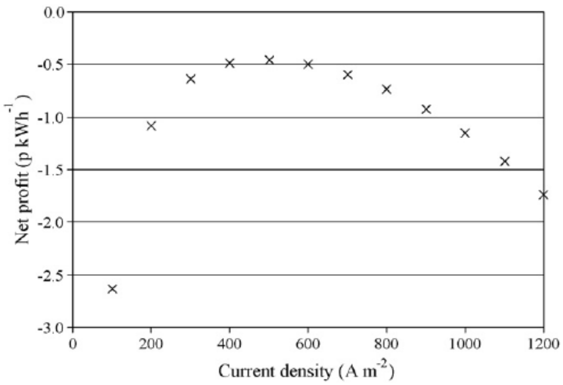

Fig. 2. Effect of current density on net profit.

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BIOGRAPHICAL NOTE

Conference presenter: Dr. Roberts is a Reader at the School of Chemical Engineering at the University of Manchester. He is an electrochemical engineer with 20 years of research experience on a wide range of technologies for energy and environmental applications. He has published more than 50 papers in international peer-reviewed journals and holds eight patents and patent applications. His research interests have covered a wide range of topics, including nanomaterials/electrocatalysis, electrokinetic treatment of contaminated land, electrosynthesis, water and waste treatment, metal recovery, fuel cells, and redox flow batteries. Dr. Roberts is a co-founder and Research Director of Arvia Technology Ltd., a successful University spin-out company that is commercializing a waste and water treatment technology. His work has been recognized through a number of awards including the European Academic Enterprise (ACES) Fast Start-up award (2009), the IET Innovation Award for Sustainability (2009), and the IChemE Water Innovation Award (2008). Dr. Roberts is a Fellow of the IChemE, a committee member and former chair of the SCI Electrochemical Technology Group, and a member of the European Federation of Chemical Engineers working party on electrochemical engineering.